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EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS FOR NEW BRUNSWICK

by

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Consulting Engineers

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EVALUATION OF
REFRACTORY GOLD PROCESSING OPTIONS
FOR NEW BRUNSWICK

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1.0 EXECUTIVE SUMMARY

This report provides an overview of the technologies available to treat refractory gold ores together with selection criteria for technologies designed to treat ores having an arseno-type refractory matrix. In addition, a bench scale test program was used as a basis for developing capital and operating cost estimates for the three most viable technologies. As a planning guide, the relative economics of selected technologies have been compared for a centralized processing facility or custom mill and a conventional integrated stand alone plant.

An evaluation format was developed to make a preliminary assessment and to select three oxidation technologies for a bench scale lab test program. The evaluation criteria were designed to determine the relative maturity of the technologies. Production cost, development status and commercial operating considerations were included as evaluation criteria. Based on a literature evaluation of all known gold oxidation technologies, biological oxidation, acid pressure oxidation, and fluid bed roasting were selected for bench scale characterization. In addition, a limited number of tests were conducted to characterize the ARSENO process.

A bulk sample of ore was obtained from the Elmtree deposit in the west Gabbro zone, at a location referred to as Trench Number 22 by Corona Corporation of Bathurst, New Brunswick.. Assays of gold, by previous trenching in this area averaged 7.54 g/mt although the sample obtained for the present evaluation averaged 5.71 g/mt. In order to assure that the oxidation rate was based on all refractory gold minerals, a high flotation recovery of gold was required. As a result, in order to achieve a gold recovery of 93% in flotation, the grade of gold in concentrate was 24.4 g/mt.

The results of the laboratory simulations can be summarized as follows:

- *Fluidized Bed Roasting* - The maximum recovery of gold was found to be 79.2% using a muffle furnace to roast the concentrate at temperatures in the range of 550 to 665°C.

- *Acidic Pressure Oxidation* - The maximum gold recovery was 96.1 % after two hours of oxidation at 200°C.

- *Bioleach Oxidation* - The maximum recovery of gold was 90.4% after 18 days with a dormant phase occurring at day 15 at a gold extraction of approximately 85 %.

In addition, a preliminary evaluation of the nitric acid catalyzed pressure leach, known as the "ARSENO Process", was designed to compare the residence time to conventional pressure oxidation methods. Overheating of the leach resulted in the formation of elemental sulphur and maximum gold recoveries of 89.7%, although the reaction conditions were not ideal.

In order to compare the preliminary economics of the three technologies for the treatment of New Brunswick's refractory gold deposits, two processing approaches were conceptually designed and costed. The base case envisaged the construction of a fully integrated flotation, oxidation and cyanidation processing facilities at a mine site. As an alternative, a custom mill designed to treat flotation concentrates from surrounding mines and/or imports from off-shore was developed.

As a basis for developing capital and annual operating cost estimates, preliminary process flow drawings were developed for both the integrated and custom mill approaches at a scale of 75,000 troy ounces of production per year. Preliminary design calculations were initially based on an integrated plant operation using an ore grade of 5.71 g Au/mt ore. The custom mill facility is derived from these estimates based on concentrate requirements at a grade of 22.5g Au/mt. Where possible, the results from the test program were used as preliminary design criteria.

Based on the results of the laboratory testing program, the required throughput capacity for the three processing options were calculated for the integrated and custom mill plants as follows:

> *Acid Pressure Leach* at 519,618 mt/year at ore of an overall gold recovery of 89.4% or 113,092 mt/year of concentrate at a gold recovery of 96.1%.
>
> *Biological Oxidation* at 540,533 mt/year of ore with an overall gold recovery of 84.1% or 117,655 mt/year of concentrate at a gold recovery of 90.4%.
>
> *Fluidized Bed Roasting* with a gold recovery of 73.7% required 658,686 mt/year ore, or 143,348 mt/year of concentrate at a gold recovery of 79.2%.

The following table summarizes the conceptual capital investment requirements and the annual operating and maintenance costs for the three technologies for both the integrated and custom mill approaches.

Capital and Annual O&M Cost Estimates
(Millions of 1988 CDN $'s)

	Acidic Pressure		*Bio-Leach*		*Fluidized Bed*	
	Capital	O&M	Capital	O&M	Capital	O&M
Integrated Plant	64.4	32.1	52.7	31.0	61.7	32.8
Custom Mill	51.5	35.6	39.4	34.5	47.6	36.9

NOTE: For comparison purposes, the annual revenue for 75,000 troy ounces of production would be $ 36 million based on a gold value of $ 400 US per ounce.

Clearly, given the grade of the Elmtree ore and the current depressed price of gold, none of the processing options evaluated yields an economic investment opportunity at the present time. However, it has been possible to perform a comparative analysis of the three options studied, in order to identify a preferred development direction.

All cases were evaluated in constant 1988 Canadian dollars, assuming 100 % equity and a fifteen year project life. In view of the negative cashflows for all of the Custom Mill cases, it was considered superfluous to introduce considerations relating to interest during construction and debt leveraging. As a result, all capital expenditures are assumed to occur in the first year of production. In addition, the sensitivity of the project to increases or decreases in capital and operating costs, as well as revenue, were evaluated over a range of +/- 30 %.

The results of the economic analysis for each option are summarized below:

Internal Rates of Return

Technology	*Custom Mill (%)*	*Integrated Plant (%)*
Bio-Oxidation	-6.88	5.41
Acidic Pressure Oxidation	-21.1	-1.49
Fluidized Bed Roaster	-8.38	5.01

In terms of economic sensitivities, it has been found that all options are relatively insensitive to capital costs due to the fact that the projected annual operating and maintenance costs represent over 50 % of the total capital investment in all cases. As well, net revenue, calculated as total revenue less total O&M expenses plus debt expense is negative. As a result, a 30 % decrease in the capital cost of a Biological Oxidation custom mill, would result in the rate of return improving from -6.88 % to -3.08 %. However, a 30 % decrease in operating costs or alternatively, a similar increase in the price of gold, would improve the rate of return on the project to +40 to +45 %. For the Fluidized Bed Roaster and Acidic Pressure Leach options, the sensitivities are equally dramatic although the magnitude of the improvement is somewhat less. In addition, it should be noted that the above return for the roaster case assumes a sale for the sulfuric acid by-product. Clearly, should the price of gold reach $ 600 per troy ounce, the economics of all options would improve, although the Biological Leach technology would remain the most attractive.

These results indicate that the Biological Leach option results in lower annual costs for both the custom mill and the integrated plant. This is partially due to the lower annual capital payment. In the case of Fluidized Bed Roasting, the major difference relates primarily to the poor gold recovery from feedstock ores or concentrates as found during the bench scale test program.

The evaluation completed to date has yielded comparative results which may be of assistance in planning further development initiatives. Clearly, aside from a sharp increase in gold prices, a successful project will require substantial improvement in the operating and capital costs required per tonne of feedstock processed. Principle areas which may warrant further consideration in an effort to optimize the choice of technology for a successful commercial venture, include the following:

o It may be beneficial to consider continious piloting of the bioleach concept in order to more precisely determine the residence time required and evaluate heat load and robustness of the process.

o It would appear that the potential for a substantial improvement in reaction rate and corresponding throughput increase may be achieved by the Arseno technology. Given the potential to substantially reduce at least the capital cost of a custom mill, further evaluation of this concept may have potential to yield a commercial project.

o In addition to the choice of oxidation technology, it is considered critical to ensure that the feedstock selected strikes the optimum balance between processing cost and grade achieved. In this regard, further study would appear to be warranted to determine the potential for overall economic improvement by optimizing the design of the primary upgrading circuit.

o The effect of feedstock variability on the stability and performance of the oxidation process is considered to be of critical importance to all technical options. For this reason, it is considered important to consider this variable in future evaluations of refractory gold processing options, particularly for a custom mill concept.

In summary, it should be recognized that between 50 and 60 % of the equivalent annual costs associated with a refractory gold processing facility, are expected to be related to feedstock acquisition. As a result, economics would dictate that every effort be made to ensure that this aspect of the project be carefully optimized. Several factors will have to be balanced in this process including, grade of ore and degree of primary upgrading. The former is a function of the deposits themselves and cannot be controlled, while the latter must be resolved through a cost benefit study. This type of analysis will be more difficult for a custom mill venture where perhaps several external interests will exercise the ultimate control over development directions. However, for a custom mill to be viable, it will be essential to have a coordinated approach to this aspect of the project. In fact, it may be that feedstock costs will have to be more a function of production cost plus agreed profit margin than it is of the price of gold, in order for a custom mill to become attractive.

2.0 INTRODUCTION

Many of the gold deposits currently being examined for development in the Province of New Brunswick are refractory in nature. Although it is generally possible to use conventional preconcentration techniques on these ores, the gold is usually found to be finely disseminated within a complex sulphide matrix (typically pyrite or arsenopyrite). As a result, the processing of these concentrates is expected to be relatively capital intensive due to the need for pretreatment to convert the matrix into a relatively porous oxide form that is amenable to conventional cyanidization techniques for the recovery of gold.

Various technology approaches are potentially applicable for the purpose of improving gold recovery from refractory ores, including the following:

o Thermal Oxidation; Roasting using rotary kilns, multiple hearth furnaces or fluidized beds.

o Pressure Oxidation; using pressurized vats to oxidize the ore in an acid solution.

o Chemical Oxidation; using strong acids or bases to accomplish the oxidation of the ore at relatively low temperatures.

o Bacterial Oxidation; using oxidizing bacteria such as thiobacillus ferroxidans and sulfolobus to degrade sulphide minerals and thereby liberate gold values.

The main difficulty rests with the task of evaluating these and other emerging and/or novel technologies against the specific characteristics of potential New Brunswick's ores.

This report provides an overview of the technologies available to treat refractory gold ores together with selection criteria for technologies designed to treat those having an arseno-type refractory matrix. In addition, a bench scale test program is used to determine capital and operating cost estimates for the three most viable technologies. The report also includes Appendices of preliminary design criteria and a separate volume that provides further details on the bench scale test program.

2.1 Project Objectives

The objectives of this study were to determine the technologies best suited to the processing of the developing refractory gold reserves of Northern New Brunswick for optimum gold extraction and to present preliminary capital and operating cost estimates for the most appropriate technologies, based on a custom mill scenario.

2.2 Study Approach

In order to address the above objectives, the study was undertaken by completion of the following tasks:

o Perform a comprehensive search of the technical literature in order to identify and assess potential refractory gold processing options that may be most suitable for New Brunswick resources.

o Select the most promising options for bench scale evaluation based on applicability to New Brunswick resources, recovery efficiency, processing economics, environmental impact and state of development.

o Procurement of a representative bulk sample of refractory ore from a known deposit in the northern portion of the Province for testing purposes

o Produce an upgraded concentrate by conventional flotation methods and document the grade/recovery relationship, as a prerequisite to bench testing of the oxidation technologies.

o Conduct a bench scale laboratory evaluation on the concentrate produced, using the most promising processing options identified during the literature review, in order to establish the ultimate gold recovery potential and to derive preliminary design and operating criteria.

o Prepare conceptual design descriptions and flowsheets for the selected processing options together with capital and operating cost estimates with an accuracy of ±30%.

o Assess the economic feasibility of a custom mill plant based on cash flow and economic sensitivities.

2.3 Previous Development Studies

Prior to this study, a number of preliminary laboratory processing studies had been conducted by Corona and the Department of Natural Resources and Energy (DNRE) on the Elmtree deposit in the time period 1986-1987. A summary of the developments to date include:

o Physical upgrading tests conducted at DNRE and Lakefield Research (LR) have indicated that the upgrading of ores by flotation may be more advantageous than gravity separation. Detailed flotation tests performed at LR have provided a basis for the design of a conceptual flotation circuit.

o Bio-oxidation tests performed on the Elmtree ore by Giant Bay Biotech Inc., under contract to Corona, have indicated that the ore can be oxidized using microorganisms, but sufficient data has not been provided to evaluate whether the technology is economically feasible.

o Roasting and cyanidization tests performed by Lakefield Research has indicated a recovery limitation for this technology. However, the roast test conditions may have favored the formation of iron arsenates and further testing will be required to determine the optimum oxygen levels and reaction temperatures.

o Detailed mineralogy and association of gold in both the Elmtree deposit and the Durham mine occurrences has been prepared by Witteck Developments under contract to DNRE. In both cases the samples were determined to be refractory due to fine grains of gold encapsulated by arsenopyrite, pyrrhotite, and pyrite. It was also observed that the ratio of arsenopyrite to pyrite and pyrrhotite varied from 1.2:1 to 2.69:1 for the Elmtree and Durham samples respectively.

2.4 The Custom Plant Concept

The concept of developing a centralized processing facility or custom mill for the recovery of gold from New Brunswick's refractory resources appears to provide a promising approach for the following reasons:

o The relatively large capital investment could be shared among several developers with processing charges based on a "user pay" formula.

o The economies of scale introduced by a large centralized concentrate processing facility would enhance the competitive position of New Brunswick developers and provide a measure of security for the industry.

o The availability of a centralized mill designed to treat concentrates produced in the Province would encourage small developers to enter the business.

In addition, optimum site selection would at once minimize processing and transportation costs to individual developers and provide the option to process concentrates produced outside of the Province if the opportunity were to arise.

A major concern with the custom plant concept is the composition and variability of feedstocks. It is anticipated that a commercial custom plant would be designed to treat concentrates of surrounding ore bodies in northern New Brunswick. Since the business development plan is beyond the scope of this study, it is assumed that the custom mill will include a comminution and beneficiation circuit in addition to oxidation, cyanidization and gold recovery units. The emphasis and scope of this study, have been placed on an evaluation of arsenopyrite oxidation technologies. In addition, conventional grinding, flotation, cyanidization, and carbon in pulp (CIP) circuits have been included for completeness in evaluating the economics of the selected technologies.

A generalized schematic for a refractory gold processing plant is presented in Figure 2.1. For a custom mill, the primary upgrading circuit may be reduced in size or eliminated, depending on the grade of available feedstocks. Following the oxidation step, the feed is treated with cyanide solution prior to recovery of the precious metals.

2.5 Resource Perspective

A number of Northern New Brunswick refractory deposits are in the early stages of development although reserve figures and ore grades have not been published. Several areas in the Upsalquitch River area and Antinouri Lake region have been staked by Lacana, Brunswick and others. In addition, refractory gold occurrences have been reported at the Durham Resources antimony mine as well as in the concentrator tailings at the Caribou base metal mine.

The intended focus of the current study has been to provide a planning tool to assist in the co-ordinated development of these deposits if and when sufficient reserves are confirmed. A custom mill concept would also be amenable to the processing of New Brunswick concentrates supplemented with offshore material should indigenous reserves fall short of requirements for a world scale processing facility. In this manner, the Province's smaller deposits could be developed without economy of scale penalties.

3.0 PROCESS TECHNOLOGY OPTIONS

The objective of this section is to provide the reader with an overview of current technical options for processing arseno-refractory ores and thereby lay a foundation for understanding the rationale utilized in evaluating the various alternatives later in this report.

As mentioned previously, the information that follows represents a synthesis of the major conclusions of studies previously undertaken on New Brunswick refractory ores as well as a comprehensive literature search. The latter was specifically designed around oxidation technologies and processing economics for ores and concentrates containing arsenic and sulphides. "Dialog Information Services Inc." drew upon a variety of databases dating from 1968 to accomplish this task.

Most of the developments in oxidation technology to treat refractory gold ore have taken place in this decade, probably due to the increase in gold prices and the fact that fewer free non-refractory deposits are available. The development of new technologies has taken place in Canada and to a lesser extent in South Africa. A summary description of the refractory gold processing technologies found in the literature is presented in this following sections.

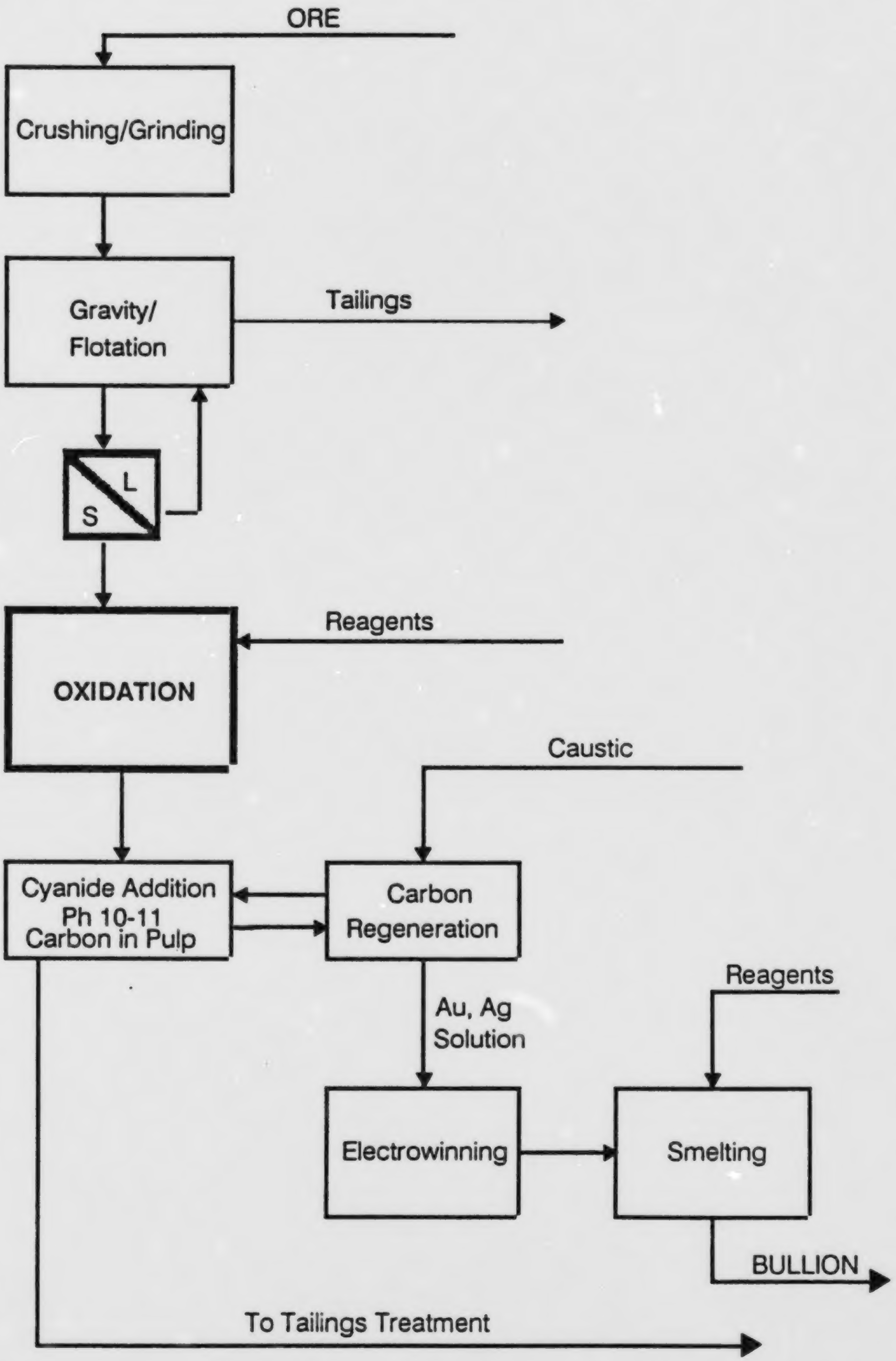

ORE
Crushing/Grinding
Gravity/
Flotation
Tailings
L
S
Reagents
OXIDATION
Caustic
Cyanide Addition
Ph 10-11
Carbon in Pulp
Carbon
Regeneration
Reagents
Au, Ag
Solution
Electrowinning
Smelting
BULLION
To Tailings Treatment

Figure 2.1 Generalized Schematic for Refractory Gold Processing

3.1 Biological Oxidation

Bioleaching technology is a relatively new concept and the majority of bench scale testing and pilot programs undertaken to date have been concentrated in Canada. Commercial plants are reported to be in operation in South Africa, but little has been published in this regard. More positively, a commercial operation is scheduled for start-up in early 1989 in Eureka County, Nevada by the Denver-based "US Gold" and several current operations are actively considering the installation of bioleach circuits as an add-on to existing sulphide flotation circuits to treat gold tailings.

Process Description

As indicated in Figure 3.1, the driving force behind the technology is the oxidation of iron and sulphur in continuously stirred tank reactors by the bacterium "Thiobacillus Ferrooxidans", or "TF". Oxidation takes place in an acidic environment (pH 1 to 2) at a temperature of 30 to 35°C in accordance with the reaction:

$$4FeS_2 + 15O_2 + 2H_2O = 2Fe_2(SO_4)_3 + 2H_2SO_4$$

Leach slurry to the tank reactors is maintained at 10 to 15% solids while the biomass concentration is in the order of 3 to 5% by volume.

Air and nutrients are continuously added to the leach circuit to maintain a high oxidation potential (so that arsenic and antimony remain in solution) and ideal conditions for regeneration of the biomass. Nutrient solutions are usually composed of ammonium sulphates required in the leach reactors to maintain biological activity at a temperature of less than 40°C (the reaction is exothermic), and in that regard, it is worthy to note that an increase in sulphur content of the feedstock will result in an increase in leach solution temperature.

System performance is sensitive to feedstock variation, in fact, each feedstock may require a specially grown TF strain to withstand specific inhibitory

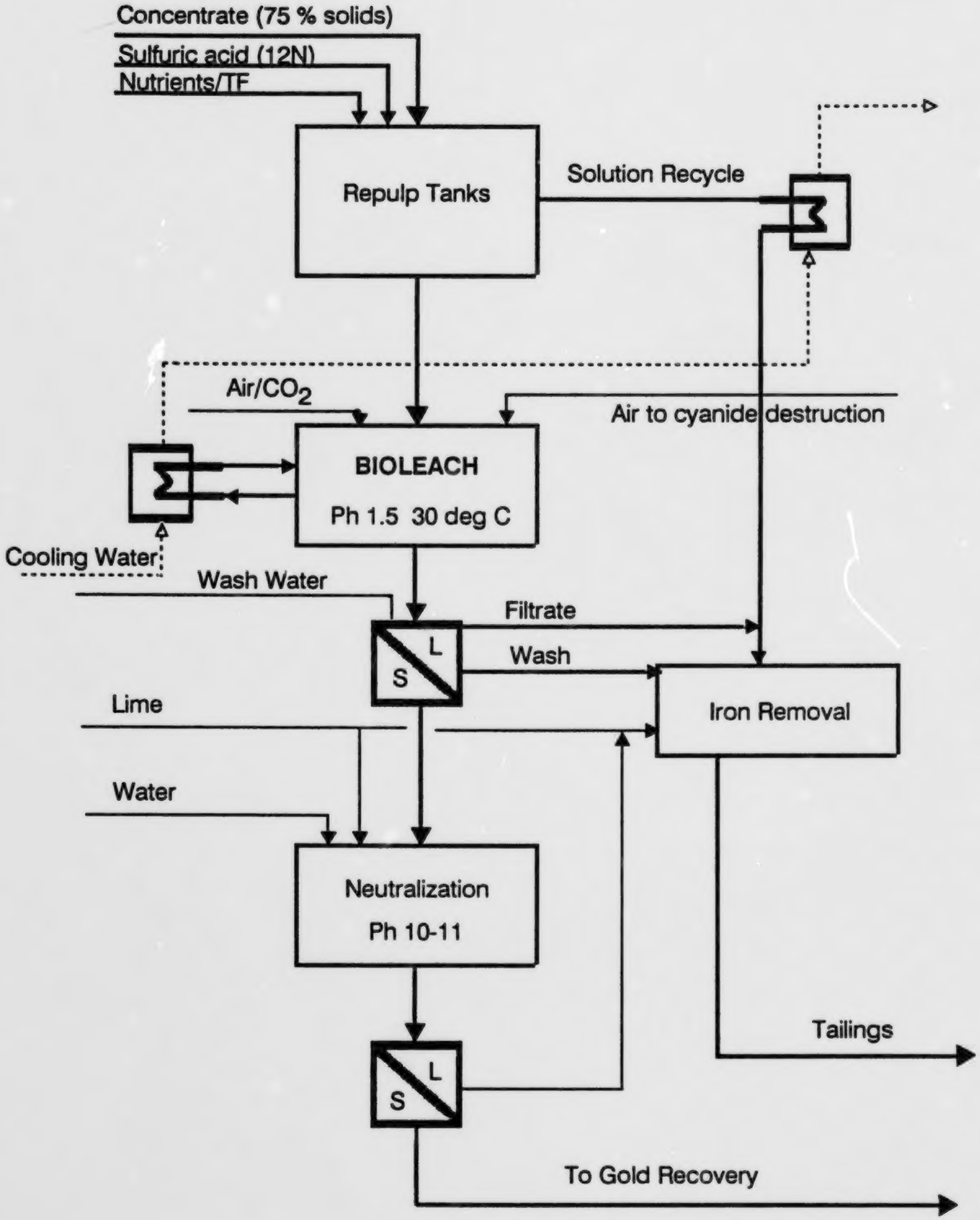

Concentrate (75 % solids)
Sulfuric acid (12N)
Nutrients/TF
Repulp Tanks
Solution Recycle
Air/CO2
Air to cyanide destruction
BIOLEACH
Ph 1.5 30 deg C
Cooling Water
Wash Water
Filtrate
Wash
L
S
Iron Removal
Lime
Water
Neutralization
Ph 10-11
L
S
Tailings
To Gold Recovery

Figure 3.1 Bioleach Process Schematic

components. Moreover, the culturing technique utilized in growing the bacteria can have a major impact on the ultimate system leaching rate. Additional concerns related to the maintenance of the biomass include:

(i) Agitation and aeration rates in relation to solids throughput are critical parameters in preventing "washout" of the biomass.

(ii) Lime used for pH control must be free of organics to avoid interference with biomass regeneration.

(iii) High concentrations of arsenic and antimony in the leach solutions inhibit the growth of TF.

(iv) Leach solutions must be maintained at a pH of greater than 1.0; this may be difficult with high sulphur ores.

(v) The bioleach process decomposes arsenopyrite much faster than pyrite; when the concentration of arsenic reaches inhibitory levels, the leach rate slows down.

There are also a number of concerns related to iron and its chemistry including:

(i) The source of water has a significant impact on iron extraction rates in that its buffering capacity or the presence of surfactant may inhibit bio-oxidation.

(ii) The precipitation of ferric arsenate by ferric sulphate in the bioleach circuit may result in excessive alkali and cyanide consumptions in the gold extraction circuit.

(iii) The iron concentration of the filtrate has to be controlled to maintain a stoichiometric ratio that will ensure concentration of arsenic.

(iv) Bioleach residues must be re-pulped and washed at a pH of 3 to prevent the precipitation of iron.

What perhaps best puts the impact of the above-noted variables into perspective is the fact that residence times in the bio-leach circuit can range from 15 to 150 hours.

Process Economics

While a comprehensive assessment of process economics is best left to Section 6, there are a few points worthy of note at this point.

(i) Capital and operating costs have been determined by Giant Bay Resources Ltd. and Wright Engineers Ltd. for a commercial plant having a capacity of 100 tpd of concentrate. Costing was based on the 1984-85 pilot program at Giant Yellowknife Mines Ltd. and was compared to both roasting and pressure oxidation.

(ii) Costing studies done by Davy McKee (Stockton), have illustrated the need for pre-concentration of ores.

(iii) A summary of commercial operating costs has been reported by Coastech Research based on pilot plant trials at Equity Silver Mines done in 1984.

(iv) Most published capital and operating cost estimate comparisons have indicated that the bio-oxidation process has the potential to be less capital and operating cost intensive than any other commercial process.

Conclusions

There are a number of gaps in the literature, including the effects of bio-oxidation on the cyanide and gold recovery circuits, the long term effects of arsenic in the gypsum tailings pond, and a lack of specific detail regarding heat loads on reactors and/or cooling requirements. Perhaps most significant in this regard is the fact that all of the ores that have been piloted to date are substantially different than the Corona deposit in terms of iron and arsenic content.

There are also a number of risks associated with this technology:

(i) Control of the biomass concentration - once lost, it may take 5-10 days to re-establish a suitable concentration.

(ii) There has been no data published to illustrate that the process is capable of remaining at a steady state for long durations with expected variations in feed composition.

(iii) There would be scale up risk associated with the design of an air injection system and the agitation of slurry in the reactors.

On the other hand, the technology has several positive features. The reactors used in the bioleach circuits can be easily fabricated and corrosion is limited to general considerations for low pH environments. New strains of bacteria are being developed to survive at elevated temperatures and it appears that the use of TF will not be subject to patent problems. Although some technical support would be required, labor skills for a bioleach operation would be minimal. Finally, and most significantly, bio-oxidation has been speculated to offer potentially lower operating and capital costs than any other oxidation technology.

3.3 Fluidized Bed Roasting

Roasting technologies have been in commercial operation for the last 40 years and there are a number of research firms familiar with fluidized bed roasting and capable of developing the process parameters required for application to specific ores. In fact, more than ten commercial fluidized bed roasters have been built to date for the treatment of refractory ores and concentrates and, as a result, the design parameters for these roasters are well defined in the literature. Patents exist on certain types of roasters and in some cases the ideal conditions for the recovery of antimony and arsenic from off-gases may be considered proprietary, although no patents exist on the process technology.

In short, this is a mature, well documented technology with a proven track record in specific applications.

Process Description

Fluidized bed roasters offer the advantages of good mixing and control of reaction temperatures so critical to the roasting of arsenic concentrates. Arsenic can encapsulate the gold in solid forms of As_2O_5 and $FeAsO_4$ if the amount of oxidation is not carefully controlled.

As illustrated in Figure 3.2, concentrate is fed to a two-stage roaster as a slurry which typically contains 70% solids. Most of the arsenic is removed in the first stage while the second is required to remove the partially decomposed pyrite. Since temperatures above 600°C result in a high formation of iron arsenates, conditions in the first stage are maintained at 500 to 550°C with an air stoichiometric ratio of 0.8 to 0.85. Temperatures in the second stage are maintained at 650 to 700°C with an air stoichiometric ratio of 1.1 to 1.2. As might be expected with a fluidized bed, feed composition and particle size are also important parameters affecting roaster performance. The operation requires a sulphur content of 15 to 20% in the feed for autogenous heat loads and silica should not be present due to associated sintering problems.

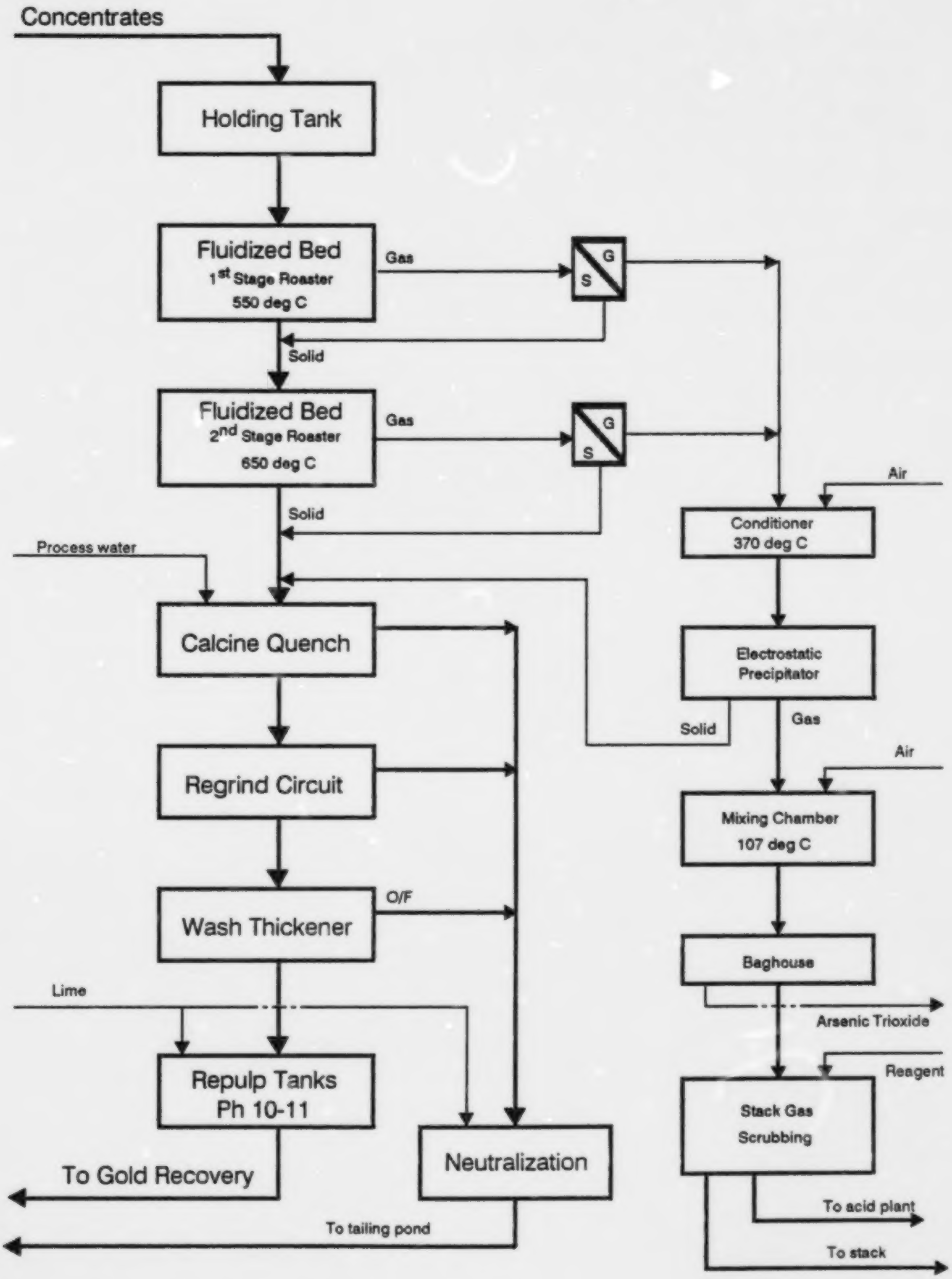

Concentrates
Holding Tank
Fluidized Bed
1st Stage Roaster
550 deg C
Gas
G
S
Solid
Fluidized Bed
2nd Stage Roaster
650 deg C
Gas
G
S
Solid
Air
Conditioner
370 deg C
Process water
Calcine Quench
Electrostatic
Precipitator
Solid
Gas
Air
Regrind Circuit
Mixing Chamber
107 deg C
O/F
Wash Thickener
Baghouse
Lime
Arsenic Trioxide
Reagent
Repulp Tanks
Ph 10-11
Stack Gas
Scrubbing
Neutralization
To Gold Recovery
To acid plant
To tailing pond
To stack

Figure 3.2 Roaster Process Schematic

Calcine dust is removed from the off-gas stream and the temperature of the remaining gas is reduced by quenching with ambient air to sublime arsenious oxide to arsenic trioxide, the latter of which takes place at a temperature of between 100 and 127°C. The acidic gas is either scrubbed and neutralized with lime or utilized in the subsequent production of sulphuric acid.

The calcine is cooled via quenching and may be reground to expose gold; fine calcine is washed in a thickener to remove sulphate. It is then slurried at an alkaline pH for treatment with cyanide. It is worthy to note that the kinetics of cyanidization have been known to be faster for calcines than non-refractory ores.

Process Economics

Operating and capital costs have been published for various commercial roasters, designed for pyrite concentrates. On a relative scale, the capital for a roaster and an acid plant has been found to be greater than a pressure leach circuit with an oxygen plant for capacities under 250 tpd of concentrate. For capacities greater than 250 tpd, the roaster and acid plant requires less capital.

While the marketing information on by-products is well defined, the economics of this process are very dependent on the day-to-day marketing possibilities, i.e. of arsenic trioxide and sulphuric acid. A few examples to illustrate:

(i) While the operating cost of a roaster with acid gas scrubbing is greater than that of an autoclave plant, the incremental cost can be more than offset if sulphuric acid market considerations are favorable.

(ii) Site requirements for land would be reduced if gypsum settling ponds were not required (i.e. if an acid plant is feasible).

(iii) Storage and/or disposal facilities are required for arsenic trioxide if it cannot be sold.

Conclusions

(i) Equipment life and reliability are influenced by high temperature acidic corrosion.

(ii) The recovery of arsenic in the off-gas is associated with two environmental problems:

o the quenching of the gas stream dilutes the sulphur dioxide content in the gas, making acid production uneconomical (Subsequent neutralization of effluent may be required).

o operation of the bag house collecting arsenic, has a high vapour pressure of arsenic trioxide, resulting in an environmentally unacceptable condition (a need to carefully control the working environment).

(iii) Highly trained operating and maintenance staff will be required in addition to technical support.

(iv) The ability to control the process at steady state has been demonstrated on a commercial scale.

3.4 Acidic Pressure Oxidation - Sherritt Process

Two commercial plants of this type have been built to date, one at San Bento Mineracao, Brazil and the other at the McLaughlin Project in California; both plants employ acid pressure leach technology on sulphide concentrates. Piloting and process development studies are also being carried out on the Porgera concentrates of New Guinea. A number of patents exist covering a range of operating temperatures and pressures, and various engineering firms have experience with the design and procurement of pressure autoclaves in the zinc and uranium industries, although experience with refractory gold ores may be limited to Sherritt Engineering. For these reasons, it is believed that conditions under the Sherritt Patent may be the most desirable for high arsenic concentrates.

Process Description

A generalized pressure leach schematic is presented in Figure 3.3. Slurry is added to a recycle acid stream and is heated in re-pulp tanks and from there is pumped to an autoclave where it is reacted with oxygen under pressure. The pressure is reduced by flash tanks and the resultant steam is either vented or utilized to preheat the slurry. Solids from the pressure leach vessels are washed in counter current decantation thickeners and, after water washing and filtration, the cake is re-pulped in an alkaline solution for cyanidization. Lime is utilized for the neutralization of wash water and the precipitation of heavy metals.

Sulphides in the concentrate are broken down to produce sulphuric acid, ferric and ferrous sulphate, and a mixture of other base metal sulphates ($MeSO_4$) in accordance with the reactions:

$$2FeS_2 + 7O_2 + 2H_2O = 2FeSO_4 + 2H_2SO_4$$
$$4FeSO_4 + 2H_2SO_4 + O_2 = 2Fe_2(SO_4)_3 + 2H_2O$$
$$MeS + 2O_2 = MeSO_4$$

These metallic sulphates (eg., potassium sulphate) will subsequently react with ferric sulphate resulting in the formation of potassium and sodium jarosites ($KFe_3(SO_4)_2(OH)_6$). Such formations result in the encapsulation of silver which may then be recovered by slaking with lime at 85°C in stirred tank reactors operating at atmospheric pressure.

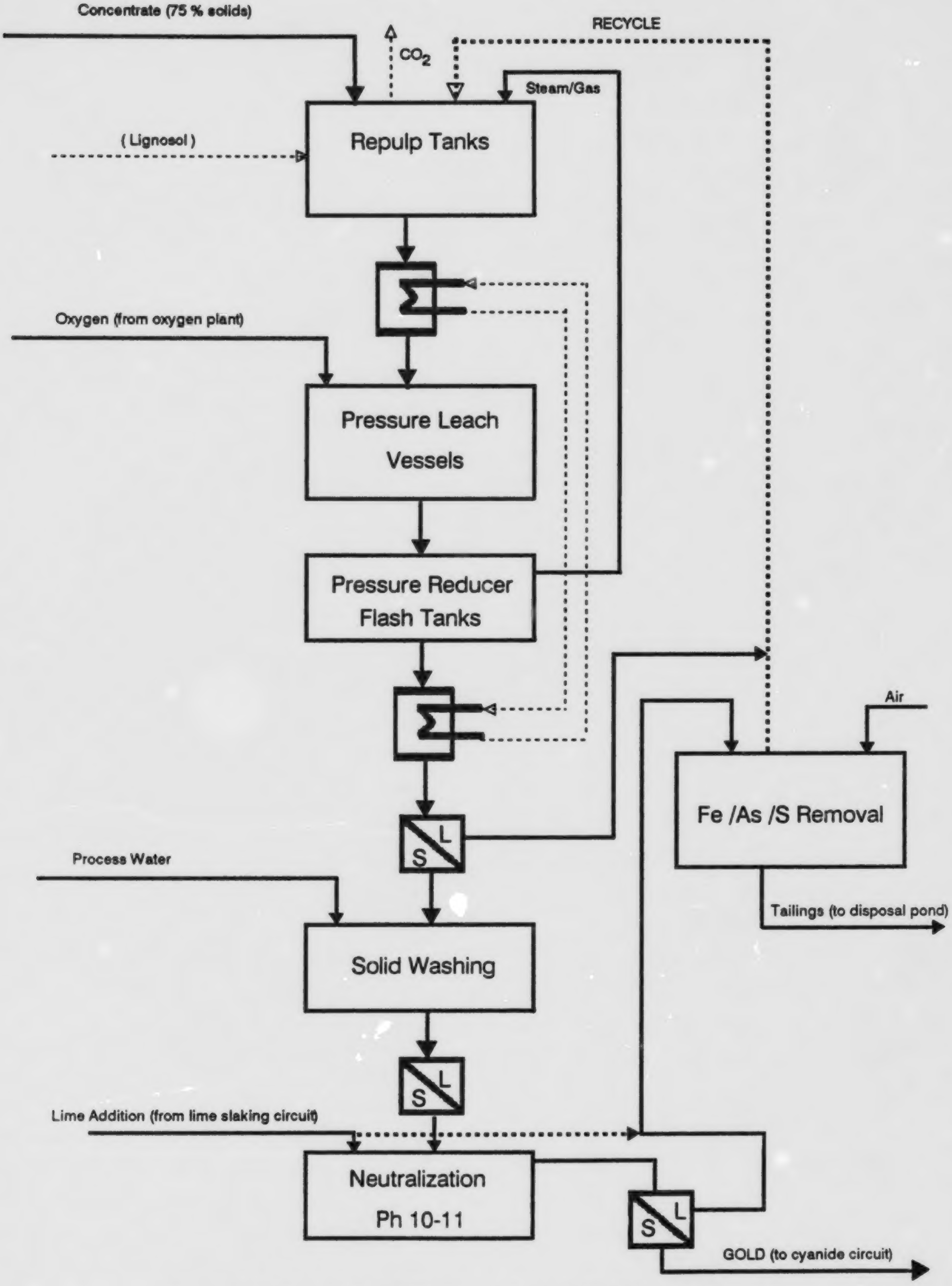

Concentrate (75 % solids)
CO2
RECYCLE
Steam/Gas
(Lignosol)
Repulp Tanks
Oxygen (from oxygen plant)
Pressure Leach Vessels
Pressure Reducer Flash Tanks
Air
S
L
Fe /As /S Removal
Process Water
Tailings (to disposal pond)
Solid Washing
S
L
Lime Addition (from lime slaking circuit)
Neutralization Ph 10-11
S
L
GOLD (to cyanide circuit)

Figure 3.3 Generalized Pressure Leach Process Schematic

Several points related to the chemistry of sulphur warrant consideration at this point:

(i) At temperatures of less than 120°C, elemental sulphur will protect sulphides from oxidation and increase the consumption of oxygen and cyanide whereas at temperatures greater than 160°C it will oxidize under pressure in the autoclave.

(ii) Complete oxidation is required to recover gold and ensure that no elemental sulphur remains.

(iii) Lignosal, a surface active reagent, must be added to the re-pulp tanks to prevent sulphur from coating the surface of sulphides.

(iv) A range of 9 to 12% sulphur is required in the reaction mixture to fuel the autoclave, which typically operates at 170 to 190°C and 1500 to 2000 Kpa with a residence time of 2 to 3 hours. The first cell of the autoclave operates at an activity of 20-30 grams per liter (gpL) H_2SO_4 and the last at 35 to 40 gpL.

(v) A sticky sulphur residue often forms near the pressure reduction zone of the flash tanks.

(vii) An autogenous heat balance can be achieved by a continuous recycle of solution, which in essence amounts to the control of sulphur content. The net sulphur requirement of the feedstock in this regard is 4 to 5%.

Process Economics

A number of economic analyses have been undertaken comparing the Sherritt Process with roasting but with conflicting results. However, it is clear that the major operating costs are associated with oxygen requirements and neutralization which reportedly represents 60% and 25% of the total operating cost respectively.

Process economics have been based on commercial operations but have not been published by Sherritt. Royalty fees may be an additional expense associated with this technology due to the rights that Sherritt maintains under Canadian patent laws.

Conclusions

The Sherritt technology can be characterized for the Corona ores using a continuous small scale autoclave since Sherritt has the required expertise in scale-up. The major risk in the design of a commercial plant is sizing of the oxygen plant and solids handling systems for pressure reduction. Caution must also be exercised in material selection to prevent fires in the autoclave and avoid high corrosion rates.

More positively, the ability to control the process at steady state has been demonstrated at the commercial scale and it may be more flexible to changes in feedstock than many other technologies.

Finally, operation of the plant will require a skilled labor force and technical support.

3.5 Catalyzed Pressure Leaching - Arseno Process

The Arseno Process was developed in 1983 by Bacon, Donaldson and Associates under a contract to Arseno Processing Ltd. and was patented in 1984. A 40 kg/hour test program was undertaken in 1986 and continuous piloting at 40 kg/hour was performed during the period September to October, 1987 utilizing low sulphide ore as feedstock. As no commercial operations to date are similar to this process it is considered to be an emerging technology.

Process Description

Concentrate is slurried with recycle liquor and fed to a continuously operating autoclave, where it is mixed with oxygen and recycled nitric oxide gas. It is then directed to a catalyzed oxygen pressure leach reactor which is cooled to remove the heat of reaction. Product is discharged to a gas separation vessel for the recovery of nitric oxide. The degassed slurry is filtered and a portion of the filtrate is used for recycle liquor; the remaining filtrate is neutralized with lime to remove heavy metals. Oxidized solids are slurried and pH is controlled via lime addition prior to cyanidization.

This process reportedly has a number of attractive features, not the least of which is that it is insensitive to feedstock sulphur concentrates ranging from 1 to 50%. Reaction rates for complete oxidation are reported to be as fast as 15 minutes. In addition, gold leaching rates during cyanidization are also reportedly fast with a retention time of 12 hours may be adequate. Moreover, iron, sulphate and arsenic remain in solution at the oxidation step and no silver is lost to jarosite, meaning that no separate processing operation is required for silver recovery.

Reaction temperatures are in the order of 80 to 100°C, oxygen pressure is between 350 and 700 Kpa, and the concentration of nitrate specified ranges from 0.2 N to 3 N (depending upon the nature of the feed).

Process Economics

Based on published information related to the pilot programs undertaken to date, it has been estimated that direct capital cost plant capable of producing 290,000 ounces of gold per year from concentrates containing 100 grams of gold per tonne would be in the order of $24.8 million. This cost is claimed to include both the hydromet circuit and the associated oxygen plant. As mentioned earlier, a major cost advantage relates to the fact that no unit operations are required for silver recovery and due to the short residence time, relatively less capital should be required for the autoclave.

Total operating costs, excluding cyanide treatment, for the above mentioned facility are claimed to be on the order of $40/tonne of concentrate, again assuming a grade of 100 grams per tonne. Thirty per cent of these costs are reportedly related to reagent consumption. It should be noted, however, that published cost estimates have been based on high grade concentrates and have not been compared to other technologies.

Conclusions

(i) The major risks in engineering and scale-up are most likely to be associated with heat load on the reactor (significant heat will have to be removed in a short time) and the recycle of nitrate gases.

(ii) The Lurgi tube reactor may be ideal for this application.

(iii) Not a commercialized process.

3.6 Alkaline Pressure Oxidation

The process flowsheet for alkaline pressure oxidation is very similar to that for acid pressure leaching, the primary difference being that sodium and oxygen (rather than recycle acid and oxygen) are utilized on the front end of the system. The concept of an alkaline-based system has been bench tested in the Soviet Union and two commercial processes have been developed: one by Equity Silver Mines Ltd. to treat arsenic and antimony in high silver/copper concentrates utilizing a two-stage pressure leach system, and the other by American Barrick Resources Corp. which featured an alkaline pressure-autoclave circuit to treat refractory carbonaceous ores.

Alkaline pressure leaching produces a residue containing iron oxide, sodium sulphate and arsenate salts. Lower oxidation temperatures are required as compared to acid pressure leaching (approximately 100°C versus 190 to 200°C) and sulphur is oxidized to a soluble sulphate.

Conclusions

The technical feasibility of utilizing alkaline pressure technology on arsenopyrite concentrates has not yet been established, but even if it were, the disposal of sodium arsenate (Na_3AsO_4) would be a major problem.

3.7 Pressure Cyanidization

The process flowsheet for pressure cyanidization is similar to that for acidic pressure leaching, the major difference being that it utilizes a tube reactor in combination with thickeners and drum filters for washing of solids from the pressure leach vessels. Bench testing of the process dates back to 1937 and has been undertaken on various ores. Pilot plant tests have been done by Consolidated Murchision on antimony concentrates.

Oxygen pressure is in the order of 5000 to 8000 Kpa at 60°C, while sodium cyanide consumption ranges from 10 to 15 kg/tonne. About 60% of the gold is reported to be recovered with a residence time of 60 to 120 minutes, but overall gold recoveries have been found to be limited for arsenic-related compounds.

3.8 Catalytic Cyanidization

Although patents exist on the utilization of potassium permanganate and peroxide as oxidizing agents during cyanidization, a fully integrated flowsheet has not been developed and very little information is available on the process.

3.9 Nitric Oxide Oxidation - Nitrox Process

While patents on the NITROX process have been filed by "Hydrochem Development Ltd.", detailed documentation on the reaction chemistry and process performance may be considered proprietary. Nonetheless, bench scale work has been completed and plans are underway to build a demonstration unit.

Process Description

As illustrated in Figure 3.4, concentrate is slurried and mixed with calcium nitrate from the off-gas scrubber, wherein:

(i) Nitric oxide is oxidized to nitrogen dioxide which is then scrubbed with water to form nitric acid.

(ii) Nitric dioxide and nitrite is scrubbed with calcium hydroxide to form calcium nitrate.

The slurry containing approximately 15% solids, then proceeds to the NITROX reactor where it is mixed with air and nitric acid under atmospheric pressure. Oxidation of arsenopyrite and pyrite is complete in one hour utilizing 12% by weight nitric acid at 80°C.

NO_x in the off-gas from the reactor is oxidized and stage scrubbed to produce nitric acid for recycle. Reaction products are filtered and the solids are re-pulped at a pH of 10 to 11 for cyanidization while the filtrate is reduced to nitrate gas, completing the recycle of nitrates.

Process Economics

While economies of the process have not been reported, it is speculated that the capital and operating costs of the nitrate recycle concept would offer potential savings over more established technologies.

Conclusions

No data is available on either of the continuous operation of the flowsheet or on the suppression of nitrates in environmental discharges.

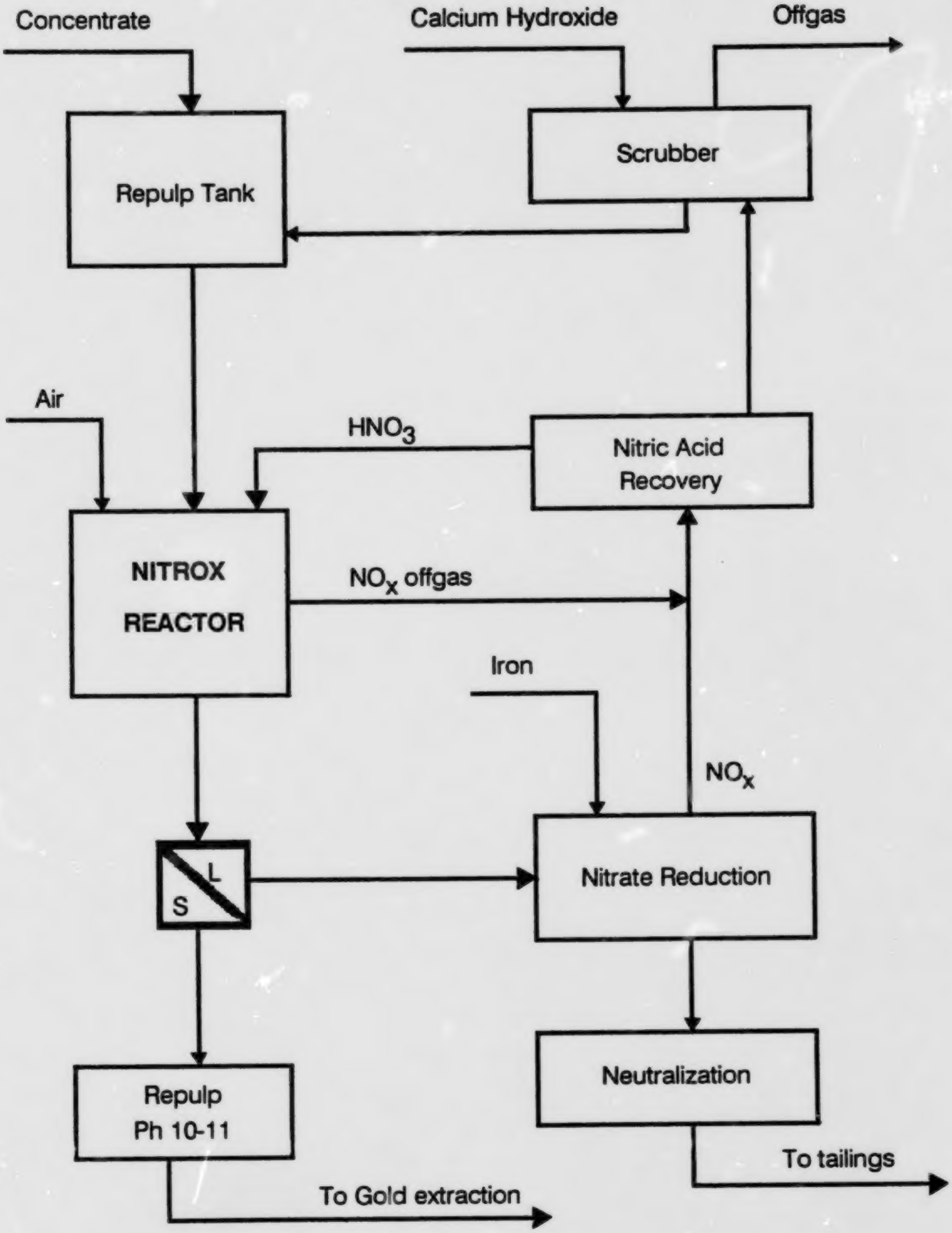

Concentrate
Calcium Hydroxide
Offgas
Repulp Tank
Scrubber
Air
HNO3
Nitric Acid Recovery
NITROX REACTOR
NOx offgas
Iron
NOx
L
S
Nitrate Reduction
Repulp
Ph 10-11
Neutralization
To tailings
To Gold extraction

Figure 3.4 Nitrox Process Schematic

Most significant, however, is the fact that the effect of the process on the cyanidization circuit has not been reported. The concern in this regard is that significant quantities of sulphate could form as a by-product of the reaction and report to the cyanide circuit, thereby effectively limiting this process to smelter feedstocks.

3.10 Chloride Hydrometallurgy

Bench scale testing of this process is ongoing at both CANMET and USBM and is centered upon the establishment of optimum process chemistry and operating conditions.

In overview, gold and silver are recovered as a by-product of the dissolution of base metal concentrates. Leaching is undertaken in a solution of hydrochloric acid and the feedstock is oxidized by the presence of ferric chloride. Acidic thiourea is utilized to recover the gold from chloride leach residues.

Three negative points related to the process within the context of this report are:

(i) The chloride leach has a lack of selectivity and dissolves most of the sulphide mineralization, producing elemental sulphur.

(ii) High levels of arsenic have a negative impact on the ferric chloride process with respect to the purification of leach solutions and electrowinning of zinc.

(iii) The economics of the ferric chloride leach are dependent upon the production of electrolytic zinc and by-products of copper and lead.

3.11 Hypochlorite Hydrometallurgy

Status

Various patents are reported in the literature on bench scale results.

Comments

The leach is not selective in the dissolution of minerals and the process is claimed to be suitable for the recovery of various sulphide metals.

3.12 High Temperature Chlorination

Status

Numerous patents on continuously operating fluidized bed chlorinators have been reported for a variety of sulphide ores. Most of the sulphides are chlorinated and the metallic chlorides are recovered from the off-gas by stage cooling. The calcines are treated for gold recovery.

Comments

Piloting of the process has been undertaken on antimony slags and concentrates and suggests that economics are dependent upon the recovery of various sulphide-based metals.

3.13 Other Oxidation Methods ... Research

Additional oxidation methods that are presently considered as fundamental research topics include:

(i) *Electrochemical Leaching*

Research is currently being carried out in the Soviet Union on gold-arsenic bearing concentrates. Preliminary calculations have suggested that electrochemical methods may be competitive with roasting.

(ii) *Microwave Irradiation*

Research is currently underway at CANMET.

4.0 EVALUATION OF OPTIONS FOR A CUSTOM MILL

Various generic studies comparing the relative production costs of oxidation technologies have recently been reported in the literature. However, the specific location of the project and the variety of refractory ore characteristics make it difficult to make general conclusions on the most viable oxidation methods. It is recognized that the characterization of the ore and the systematic testing of oxidation technologies are essential in determining the technical and economic feasibility of processing options.

An evaluation format was designed to make a preliminary assessment and to select three oxidation technologies for a bench scale lab test program. The evaluation criteria were designed to determine the relative maturity of the oxidation technologies. Production cost, development status and commercial operating considerations are included as evaluation criteria.

The objective of the preliminary evaluation was to identify the strengths and weaknesses of each oxidation technology for a custom mill. Published literature and general knowledge were utilized to make an unbiased assessment based on the evaluation criteria outlined in Table 4.1. A marking scheme was included to place weighted averages on each category. The evaluation was based heavily on the development status of the technology and the availability of information on economics, by-product markets, production costs, and environmental impacts.

A number of factors must be considered in selecting a processing technology specific to a given feedstock; however, in the case of the custom mill concept the following definitions are considered preliminary operating objectives and will provide a basis for the present evaluation:

Feedstock

The selection of oxidation technology is specific to arseno-refractory ores. Therefore, the ore types being treated by a custom mill must be similar to the Elmtree deposit. Ores that are classified as having gold "lock-up" with tellurides stibnite, galena, copper minerals, carbonaceous materials, and/or tellurides, would not be appropriate for treatment by the proposed custom mill. It is possible that supplementary feedstocks of arseno-type ores could be imported to the custom mill from various locations, but the mineralogical and oxidation characteristics would need to be similar to the Elmtree deposit.

Technology Integration

The oxidation technology must be capable of treating concentrates and producing oxidized solids amenable to cyanidization. It is evident that the primary upgrading of the ore by flotation and/or gravity separation would be required to improve the overall economics of the custom mill. Grinding of the ore to achieve a fine particle size for oxidation would have a minor impact on production cost. Due to the nature of the feedstocks being considered, the oxidation technology must be capable of treating high concentrations of arsenic, sulphur, and iron. In addition, the removal of these elements in their oxidized state is necessary to prevent high reagent consumption in cyanidization. Thus, the selected oxidation technology must be complementary to the preconcentration and cyanidization steps.

Table 4.1

Process Evaluation Criteria for Oxidation Technologies

DEVELOPMENT STATUS OF TECHNOLOGY (40%)

- o Support of process development (30%)
- o Availability of technical support (25%)
- o Commercial applications (15%)
- o Engineering scale-up data (15%)
- o Engineering management (10%)
- o Development requirements (5%)

ECONOMIC & INVESTMENT CRITERIA (35%)

- o Process economics (30%)
- o Project sensitivities (25%)
- o Economic working life (15%)
- o Additional expenditures required (15%)
- o By-product marketing potential (10%)
- o Construction schedule (5%)

CONSIDERATIONS FOR PLANT OPERATION (20%)

- o Environmental management (30%)
- o Steady state operation (25%)
- o Plant availability (10%)
- o Maintenance requirements (10%)
- o Raw materials (10%)
- o Labor skills (5%)
- o Health & safety (5%)

TECHNICAL FEASIBILITY OF FLOWSHEET (5%)

- o Operating variables (40%)
- o Mass & energy balances (30%)
- o Throughput flexibility (20%)
- o Process control (10%)

Commercialization

The development of a custom mill is being considered, based on the potential commercial benefit for facilitating the economic development of a number of small deposits. Therefore, the selection of an oxidation technology must be based on proven technology which offers an acceptable degree of operational and technical risk.

4.1 Evaluation Criteria

The following subsections provide a brief explanation of the criteria utilized in evaluating each of the technologies as well as an indication of their relative contribution (in terms of points) to the development of a final score.

4.1.1 Development Status of Process Technology

Development status measures the maturity of a particular technology and the associated availability of engineering design criteria. Technology on the leading edge of commercialization must be evaluated in the light of economic, market, and environmental information, while consideration of the evolution of a technology to commercialization, must take into account the degree of confidence in engineering scale-up and design. Technologies having commercial applications received a relatively high rating with consideration being given to feedstock type, plant location, climate, and by-product market location. The section rating accounts for 40% of the technology evaluation and is inversely proportional to the technological development required for commercialization.

4.1.2 Economics and Investment Criteria

The evaluation of the economic viability of a technology is based on the production cost data, the return investment, and the project sensitivities.

Production cost for the evaluation includes operating and capital, equipment replacement, process development, and construction. In most cases, the product cost data has been reported in the literature as conceptual costing, and is not based on actual costs of pilot or commercial operations.

The economics and investment criteria provides an assessment of the economics based on a fully integrated gold process plant, including feedstock upgrading and cyanidization. The assessment (rating) for this section makes up 35% of the technology evaluation and is proportional to the economic feasibility.

4.1.3 Considerations for Plant Operation

Two major factors to consider in the management of an operating facility are environmental impact and steady state plant operation. Environmental impact includes the assurance that environmental standards for effluent discharge, solid waste disposal, and reclamation can be achieved. In addition to tailings management, the oxidation technology may not be considered viable if the plant is faced with significant down time as a result of mechanical failure or operating complexity. Therefore, the assessment is based on the ability to control the process and operate at steady state conditions (accounts for 20% of the technology evaluation).

4.1.4 Technical Operating Feasibility of Flowsheet

Technical competence is required to maintain an efficient commercial operation. Thus, the technical ability to maintain an operating plant will include the knowledge of flowsheet performance with respect to the flexibility of operating variables and feedstock characteristics. These criteria were evaluated utilizing published data related to the appraisal of continuously operating technologies.

4.2 Summary of Evaluation Results

The overall evaluation of the technologies described in Section 3 according to the previously described criteria, is shown in Table 4.2. The detailed evaluation results by subsection are presented in Tables 4.3A to 4.3D. The final score represents the relative development potential of each technology based on present day status. A summary of strengths and weaknesses of each technology is presented in Tables 4.4A to 4.4C.

In summary, the processes judged to be most appropriate for the prefeasibility test program on New Brunswick ores, were biological oxidation, roasting, and acidic pressure oxidation. Other processes considered to be specific to certain ores included all the chlorine technologies, alkaline pressure oxidation (commercially proven for carbonaceous ores) and pressure and catalyzed cyanidization (commercially unproven). Technologies that have been specifically designed for the treatment of arseno-type ores and are reportedly to be demonstrated commercially in the near future include the ARSENO process and the NITROX process. These processes may warrant further consideration as more detailed performance information becomes available.

It is anticipated that the present day status of processing technologies will change rapidly within the next few years. New technologies will emerge, and developing technologies may be commercially proven. However,for the custom mill concept, the present day relative comparison of development potentials for each oxidation technology is as follows:

Technology	Relative Development Potential
Fluidized Bed Roasting	1.0
Acidic Pressure Oxidation	0.96
Biological Oxidation	0.84
Arseno-catalyzed Pressure Oxidation	0.52
NITROX-Nitric Acid Oxidation	0.39

TABLE 4.2

SUMMARY OF PROCESS TECHNOLOGY EVALUATION

PROJECT: Refractory Gold Oxidation Technology
DATE: October 16,1988

EVALUATION CRITERIA	PROCESS TECHNOLOGY											
	BIO-LEAC OXIDATIO	ROAST OXID.	ACIDIC PRESSURE OXID.	ARSENO CATALYST PRESSURE	ALKALINE PRESSURE OXID.	CYANIDE PRESSURE	CATALYZE CYANIDE	NITROX NITRIC ACID	CHLOR. HYDRO	HYPO-CHLORIDE LEACH	HIGH TEMP. CHLORIDE	OTHER
1.0 DEVELOPMENT STATUS OF PROCESS TECHNOLOGY (40%)												
TOTAL :	51.5	85	75	36	19.5	16	13	29.5	19.5	9	4	3.5
Section Score Based on 40%	20.6	34	30	14.4	7.8	6.4	5.2	11.8	7.8	3.6	1.6	1.4
2.0 ECONOMIC AND INVESTMENT CRITERIA (35%)												
TOTAL:	61	56	56	31	11	11.5	5	16.5	12.5	5	3	0
Section Score Based on 35%	21.35	19.6	19.6	10.85	3.85	4.025	1.75	5.775	4.375	1.75	1.05	0
3.0 CONSIDREATIONS FOR PLANT OPERATION (20%)												
TOTAL:	64	50	59	40.5	23	27	7	38.5	16	15	9	0
Section Score Based on 20%	12.8	10	11.8	8.1	4.6	5.4	1.4	7.7	3.2	3	1.8	0
4.0 TECHNICAL OPERATING FEASIBILITY OF FLOWSHEET												
TOTAL:	37	43	48	23	1	1	0	20	1.5	1	.5	0
Section Score Based on 5%	1.85	3.15	2.4	1.15	.05	.05	0	1	.075	.05	.025	0
TOTAL SCORE BASED ON 10	56.6	66.7	63.8	34.5	16.3	15.8	8.3	26.2	15.4	8.4	4.4	1.4

TABLE 4.3A

EVALUATION OF PROCESS TECHNOLOGY

PROJECT: Refractory Gold Oxidation Technology
DATE: October 16,1988

EVALUATION CRITERIA	RATING GUIDELINES		PROCESS TECHNOLOGY BIO-LEAC OXIDATIO	ROAST OXID.	ACIDIC PRESSURE OXID.	ARSENO CATALYST PRESSURE	ALKALINE PRESSURE OXID.	CYANIDE PRESSURE	CATALYZE CYANIDE	NITROX NITRIC ACID	CHLOR. HYDRO	HYPO-CHLORIDE LEACH	HIGH TEMP. CHLORIDE	OTHER
1.0 DEVELOPMENT STATUS OF PROCESS TECHNOLOGY (40%)														
1.1 Support of process development	:no information or data	0 to 5	15	23	20	10	4	4	4	6	3	2	1	1
	:degree of confidence	6 to 25												
	:proven to meet specifications	26 to 30												
1.2 Availability of technical support	:not available	0 to 5	15	23	21	16	9	9	8	14	12	6	2	2
	:technology is proprietary	6 to 10												
	:infor. available for bench test	11 to 15												
	:info. available for piloting	16 to 20												
	:info. available for commercial	21 to 25												
1.3 Commercial applications	:no commercial application	0 to 5	6	14	11	2	2	1	0	2	1	0	0	0
	:limited number	6 to 10												
	:similar applications	11 to 15												
1.4 Engineering scale-up data	:high risk	0 to 4	7	11	9	3	1	0	0	3	.5	0	0	0
	:extent of risk	5 to 10												
	:limited risk	11 to 15												
1.5 Engineering management	:no experience with technology	0 to 4	5	9	9	2	2	0	0	2	1.5	0	0	0
	:degree of confidence	5 to 7												
	:similar experience	8 to 10												
1.6 Development requirements	:bench,pilot,commercial	0 to 5	3.5	5	5	3	1.5	2	1	2.5	1.5	1	1	.5
		TOTAL :	51.5	85	75	36	19.5	16	13	29.5	19.5	9	4	3.5
		Section Score Based on 40%	20.6	34	30	14.4	7.8	6.4	5.2	11.8	7.8	3.6	1.6	1.4

TABLE 4.3B

EVALUATION OF PROCESS TECHNOLOGY

PROJECT: Refractory Gold Oxidation Technology
DATE: October 16,1988

EVALUATION CRITERIA	RATING GUIDELINES		PROCESS TECHNOLOGY											
			BIO-LEAC OXIDATIO	ROAST OXID.	ACIDIC PRESSURE OXID.	ARSENO CATALYST PRESSURE	ALKALINE PRESSURE OXID.	CYANIDE PRESSURE	CATALYZE CYANIDE	NITROX NITRIC ACID	CHLOR. HYDRO	HYPO-CHLORIDE LEACH	HIGH TEMP. CHLORIDE	OTHER
2.0 ECONOMIC AND INVESTMENT CRITERIA (35%)														
2.1 Process economics	:no information on costs :not considered a good investment :degree of confidence	0 to 5 6 to 10 11 to 30	15	24	20	15	2	3.5	1	2	2	0	0	0
2.2 Project sensitivities	:highly sensitive or unknown :degree of sensitivity :low sensitivity	0 to 5 6 to 15 16 to 25	16	6	10	2	1	1.5	0	2	1.5	0	0	0
2.3 Economic working life	:limited life :degree of economic life :no limitations to life span	0 to 4 5 to 10 11 to 15	12	6	9	3.5	3	3.5	1	3.5	2	1	.5	0
2.4 Additional expenditures	:major additional expenses :degree of additional expenses :limited additional expense	0 to 3 4 to 10 11 to 15	10	13	11	3.5	1	1	0	3	.5	0	0	0
2.5 By-product potential	:no potential by-products :degree of feasibility :well defined and market is stable	0 to 3 4 to 7 8 to 10	3	5	3	3	2	0	0	3	4	1	1	0
2.6 Construction schedule	:extent of time required	0 to 5	5	2	3	4	2	2	3	3	2.5	3	1.5	0
		TOTAL:	61	56	56	31	11	11.5	5	16.5	12.5	5	3	0
		Section Score Based on 35%	21.35	19.6	19.6	10.85	3.85	4.025	1.75	5.775	4.375	1.75	1.05	0

TABLE 4.3C

EVALUATION OF PROCESS TECHNOLOGY

PROJECT: Refractory Gold Oxidation Technology
DATE: October 16,1988

EVALUATION CRITERIA	RATING GUIDELINES		PROCESS TECHNOLOGY											
			BIO-LEAC OXIDATIO	ROAST OXID.	ACIDIC PRESSURE OXID.	ARSENO CATALYST PRESSURE	ALKALINE PRESSURE OXID.	CYANIDE PRESSURE	CATALYZE CYANIDE	NITROX NITRIC ACID	CHLOR. HYDRO	HYPO-CHLORIDE LEACH	HIGH TEMP. CHLORIDE	OTHER
3.0 CONSIDREATIONS FOR PLANT OPERATION (20%)														
3.1 Environmental Management	:no envir. protection considered :limited tech. to support man. :security of envir. protection	0 to 5 6 to 15 16 to 30	20	12	18	8	2	2	0	8	1	1	0	0
3.2 Steady state operation	:not proven to run at s.s. :tested to run at s.s. :degree of automation required	0 to 5 6 to 10 11 to 25	12	19	15	3	0	2	0	3	0	0	0	0
3.3 Plant availability	:high down-time :impact of avail. on operation :limited down-time	0 to 3 4 to 7 8 to 10	5	4	7	7	3	3	0	6	4	3	2	0
3.4 Maintenance	:degree of maintenance required :limited maintenance	0 to 5 6 to 10	10	3	6	8	6	6.5	0	8	4	4	2	0
3.5 Raw materials	:limited availability :locally available	0 to 5 6 to 10	8	9	7	7	5	5	0	6	2	2	1	0
3.6 Labor skills	:measure of training required	0 to 5	5	2	3.5	4	4	5	4	4	3	3	2	0
3.7 Health and safety	:degree of precautions required	0 to 5	4	1	2.5	3.5	3	3.5	3	3.5	2	2	2	0
		TOTAL:	64	50	59	40.5	23	27	7	38.5	16	15	9	0
		Section Score Based on 20%	12.8	10	11.8	8.1	4.6	5.4	1.4	7.7	3.2	3	1.8	0

TABLE 4.3D

EVALUATION OF PROCESS TECHNOLOGY

PROJECT: Refractory Gold Oxidation Technology
DATE: October 16,1988

EVALUATION CRITERIA	RATING GUIDELINES		PROCESS TECHNOLOGY											
			BIO-LEAC OXIDATIO	ROAST OXID.	ACIDIC PRESSURE OXID.	ARSENO CATALYST PRESSURE	ALKALINE PRESSURE OXID.	CYANIDE PRESSURE	CATALYZE CYANIDE	NITROX NITRIC ACID	CHLOR. HYDRO	HYPO-CHLORIDE LEACH	HIGH TEMP. CHLORIDE	OTHER
4.0 TECHNICAL OPERATING FEASIBILITY OF FLOWSHEET														
4.1 Operating Variables	:correlation unknown	0 to 5	20	30	15	5	0	0	0	5	0	0	0	0
	:degree of variance tested	6 to 35												
	:operating range well defined	36 to 40												
4.2 Mass and energy balance	:not defined	0 to 5	5	15	7	7	0	0	0	4	0	0	0	0
	:degree of variation/confidence	6 to 30												
4.3 Throughput flexibility	:variation has a large effect	0 to 5	7	10	18	5	0	0	0	5	0	0	0	0
	:degree of tolerance on variation	6 to 20												
4.4 Process control	:not defined	0 to 3	5	8	8	6	1	1	0	6	1.5	1	.5	0
	:degree of confidence in control	4 to 10												
	TOTAL:		37	63	48	23	1	1	0	20	1.5	1	.5	0
	Section Score Based on 5%		1.85	3.15	2.4	1.15	.05	.05	0	1	.075	.05	.025	0

Table 4.4A

Summary of Strengths and Weaknesses for Oxidation Technologies

TECHNOLOGY	STRENGTHS	WEAKNESSES
Biological Oxidation	o Potential savings of operating and capital costs o Mild reaction conditions. o Most engineering & development expertise in Canada. o Strong industry interest. o Extensive piloting experience.	o Little published commercial experience. o Sensitivity to variations in feedstock. o Bioleach rate inhibited by arsenic. o Uncertainty of engineering (mixing & oxygen) transfer). o Residence times can be high. o Downtime extensive if biomass lost. o High reagent costs for lime.
Fluidized Bed Roasting	o Well established commercial process. o Good mixing & control of temperatures. o Low operating cost if market for acid. o Autogenous heat balance with 15-20% S. o Process technology not patentable.	o Sensitive to feedstock variation. o High capital and operating costs. o If no acid plant, waste disposal a concern. o Arsenic poses health & safety concern. o Plant reliability/ maintenance. o Gold recovery may be limited.

Table 4.4B

Summary of Strengths and Weaknesses for Oxidation Technologies

TECHNOLOGY	STRENGTHS	WEAKNESSES
Acidic Pressure Oxidation - Sherritt Process	o Commercial plants established. o High gold recovery. o Autogenous heat balance with 4-5% S. o Economics of process well defined. o No major environmental concerns.	o High capital & operating costs. o Process is patented. o Somewhat sensitive to feedstock. o Materials of construction critical. o Ag recovery requires additional processing. o High reagent costs for oxygen and lime.
Catalyzed Pressure Leaching - Arseno Process	o Designed for high arsenic. o Short reaction time. o Lower capital cost. o Continuous pilot plant available. o No separate Ag recovery required.	o Details of process control uncertain. o Nitrate disposal implications. o No commercial operations. o Engineering and scale-up concerns.
Alkaline Pressure Oxidation	o Lower temperature reaction. o Less corrosion risk. o Established chemistry.	o Disposal of sodium arsenate in an environmental concern. o Feasibility not demonstrated. o Not specific to arseno-type ores.
Pressure Cyanidization	o Piloted on antimony arsenic conc. o No oxidation step required.	o Low gold recovery with arsenic conc. o High consumption of cyanide. o Not commercially proven. o Limited information available.

Table 4.4C

Summary of Strengths and Weaknesses for Oxidation Technologies

TECHNOLOGY	STRENGTHS	WEAKNESSES
Catalytic Cyanidization	o No oxidation step required. o Potentially faster kinetics.	o Chemistry not fully understood. o Cyanide consumption high. o No flowsheet developed.
Nitric Acid Oxidation - Nitrox Process	o Lower temperatures & pressures. o Oxidizing agent recycled, reducing reagent costs.	o Capital & operating costs not published. o Process is patented. o Not continuously piloted. o Nitrate disposal a concern. o Sulphur may be high in oxidized product.
Chloride Treatment	o For treatment of base metal sulphide conc. o Some concepts have been piloted.	o Chlorination is non-selective. o High reagent cost. o Economics related to base metals recovery. o Commercial viability unproven.

5.0 BENCH SCALE PROCESS EVALUATION

A bench scale program was designed to undertake a preliminary assessment of the applicability of the selected oxidation technologies using concentrates produced from the Elmtree ore. Based on the literature evaluation of the oxidation technologies, biological oxidation, acid pressure oxidation, and roasting were selected for bench scale characterization. In addition, a limited number of tests were conducted to characterize the arseno process. Lakefield Research of Lakefield, Ontario was subcontracted to conduct the lab program. A detailed compilation of the findings of the laboratory program is contained in Volume II of this report.

5.1 Sample Acquisition

The acquisition of the sample was subcontracted to Three-D Geoconsultants Limited of Fredericton, New Brunswick. The selection of sampling location to ensure that the sample was unoxidized and representative of the deposit, was coordinated with the geological staff of Corona Corporation of Bathurst, New Brunswick. An description of the geology and mineralogy of the deposit has been prepared by Three-D Geoconsultants Ltd. and is presented in Appendix A.

The bulk sample was taken from the Elmtree deposit in the west Gabbro zone, at a location referred to as Trench Number 22 by Corona. Assays of gold, by previous trenching in this area averaged 7.54 g/mt. To ensure minimum oxidation of the sample, the trench was blasted in two lifts. The first lift measured 3 meters by 1 meter and 1 meter deep and was intended to remove highly oxidized surface material. A second lift was blasted for bulk sampling, having an area of 2.5 metres by 0.6 metres and 0.75 metres in depth. In total, approximately 1,095 kilograms of ore were recovered and placed in four sealed drums after sampling. Analysis of the samples showed an average assay for each drum as follows (assays are expressed in grams of gold per metric tonne of ore):

DRUM #1	6.24 g/mt Au
DRUM #2	5.55 g/mt Au
DRUM #3	8.57 g/mt Au
DRUM #4	5.15 g/mt Au

Drums #1 and #4, having intermediate gold content, were sent to Lakefield Research for use in the bench scale simulations.

In spite of the precautions taken, surface oxidation was observed on the samples. This is thought to be due to the fracture planes in the bedrock which has affected the deposit to depths of several tens of metres. At the time of sampling, it was not known how this oxidation would affect the gold recovery. It was later found that direct gold extraction from the samples resulted in poor recoveries similar to earlier findings using unoxidized core samples and confirming the refractory nature of the ore.

5.2 Laboratory Test Results

A bench scale test program was designed to include the characterization of oxidation technologies and provide data for the preliminary sizing of process equipment. Conventional operating process variables were used for each technology and conditions were not necessarily optimized due to the budget constraints. Emphasis was placed on the degree of oxidation per unit of time and the relationship between degree of oxidation and gold recoveries. Details of the test procedures and results are discussed in Volume II of this report, entitled "An Investigation of the Recovery of Gold from Elmtree Ore Samples", prepared by Lakefield Research.

5.2.1 Ore Characteristics

Mineralogical examination of the ore indicated that pyrrhotite was approximately 95% replaced by marcasite (a secondary oxidation product). This is significant since pyrrhotite is oxidized more readily than other forms of iron sulphide. There was no evidence of arsenopyrite or pyrite oxidation, and the ore was classified as a refractory arsenopyrite. The degree of sample oxidation could not be measured.

Oxidation of the ore may have a direct impact on the sizing of an oxidation circuit. Theoretically, the time required for the oxidation process would be less if the feedstock has a certain degree of preoxidation. However, if gold extraction from naturally oxidized ore is similar to unoxidized ore, it would not have an impact on the oxidation circuit sizing. After discussion with officials of the Department of Natural Resources and Energy, it was decided that the small portion of naturally oxidized ore would not be removed from the bulk sample and that the rate of oxidation for each technology would be compared on a relative basis.

Semi-quantitative spectrographic analysis of the ore showed a significant concentration of alkali and alkaline earth elements such as calcium, sodium, magnesium and aluminum. The Bulk analysis of the ore received by Lakefield Research is summarized in Table 5.1.

The recovery of gold by direct cyanidization of a portion of the bulk sample was found to be similar to previous tests performed on Elmtree core samples. Cyanidization of the ore ground to 85% less than 200 mesh, was successful in extracting 33% of the gold after 48 hours. For economic reasons, cyanidization at a finer grind was not attempted on the ore. It is believed that the economics of fine grind would only be supported by preconcentration, followed by a fine regrind. Therefore, an ultra-fine grind was tested on flotation concentrate.

5.2.2 Batch Flotation

The conditions for flotation were selected from previous test results completed by Lakefield Research on core samples from the Elmtree deposit. Initial testing indicated that the ore flotation characteristics were somewhat

Table 5.1

Analysis of Ore - Elmtree Deposit

5.71 ± 0.22 g/t gold
< 2.0 g/t silver
1.73% arsenic
4.02% sulphur
10.2% iron

[10-100%] silica
[1-10%] magnesium, sodium
[0.3-3%] aluminum, calcium

[ ] semi-quantitative analysis

different than those from previous testing. Test conditions were varied to improve gold recoveries and to achieve target grade concentrate in the range of 25 to 30 g/mt Au. The overall grade-recovery relationship of a rougher flotation is illustrated by Figure 1, Page 8 of Volume II. It should be noted that improved performance should be expected from a complete rougher, cleaner, scavenger flotation circuit; however identification of the technical/economic optimum was beyond the scope of the present investigation.

The concentrate for the oxidation testing was produced by batch flotation methods. The grade of the bulk rougher flotation concentrate was 22.5 g Au/mt at 91.7% recovery of gold. Based on the flotation test results, the range of concentrate specifications for preliminary design purposes is presented in Table 5.2.

It has been found that the sulphur content of low grade concentrate may be to low to satisfy autogeneous heat loads. In addition, a limitation to increasing the sulphur and gold content of concentrates relates to the absolute gold recovery. However, the shortfall in terms of heat load is not considered crucial in that a relatively minor quantity of auxiliary heat would have to be supplied by a supplementary fuel source. The cost associated with this requirement would be minor in comparison to the cost of concentrate for the custom mill.

For chemical processing methods, the ratio of iron to arsenic may be low, compared to commercial operations. Thus, the formation of stable iron-arsenic complexes in liquid effluents from oxidation may require additional iron to satisfy the stoichiometry.

Mineralogical analyses of flotation tailings have indicated the presence of marcasite (FeS_2) and trace amounts of arsenopyrite, pyrite and pyrrhotite. The iron bearing minerals are generally entrained in gangue minerals, such as silica. In addition to a primary grind of 85 to 90% -200 mesh, a regrind may be required in order to improve on gold recoveries. This was illustrated by the preliminary regrind tests performed on tailings, which improved the gold recovery by an additional 2%.

Table 5.2

Bulk Flotation Design Criteria - Rougher Flotation

Batch Flotation Results

	MINIMUM GRADE	MAXIMUM GRADE
Gold (g/mt Au)	22.5	35.0
Arsenic (% As)	6.44	11.88
Sulphur (% S)	11.0	17.34
Iron (% Fe)	23.9	(32-38)
Overall Gold Recovery (%)	94	83
Ratio Au/S	2.0	2.0
Ratio Fe/As	3.7	(2.7-3.2)

() Estimated Value

Direct cyanidization of flotation concentrate ground to 97% minus 400 mesh(38 micron) resulted in a gold recovery of 40.9% after 48 hours. An additional test using an ultra-fine grind of flotation concentrate to 92% minus 11 micron did not result in improved recovery. Test results indicate that oxidation of the ore will be required to expose gold for cyanidization and therefore improve overall recoveries. This is due to the ultra-fine grains of gold occurring in a solid solution of coarse arsenopyrite crystals. As discussed earlier, arsenopyrite was not found to be oxidized.

5.2.3 Fluidized Bed Roasting

The commercial technology for fluidized bed roasting of arsenic concentrates has been known for over 50 years. Typically, with this technology, the recovery of gold by cyanidization is directly related to the efficiency of sulphur and arsenic removal. Previous tests performed on the Elmtree deposit indicated a limited gold recovery in the range of 75 to 81% with calcine residues of 1.7 to 3.4% As and 0.34 to 8.0% S.

The analysis of the ore used by Lakefield Research for the bench test program is summarized in Table 5.1. The recovery of gold by direct cyanidization of a portion of the bulk sample was found to be similar to previous tests performed on Elmtree core samples. Cyanidization of the ore ground to 85% less than 200 mesh, was successful in extracting 33% of the gold after 48 hours. Due to constraints of the test program, cyanidization at a finer grind was not attempted on the ore. However, it is believed that the economics of a fine grind would only be supported at temperatures in the range of 550 to 665°C. Following high temperature oxidation, the weight loss of the solids was found to be 16% of roaster feed weight. For cyanidization of the oxidation residue, the consumption of sodium cyanide and calcium oxide is 2.42 kg/mt and 11.66 kg/mt, respectively, for a 24 hour leach time.

Table 5.3

Test Conditions for Roasting Oxidation and Carbon in Leach Gold Recovery

Temperature		580-665°C
Air Ratio		(Air not controlled)
Residence Time	First Stage	30 min.
	Second Stage	60-150 min.
Particle Size		97%, -400 mesh
Cyanidization Pulp Density		30% solids
Cyanide Concentration		2.0 g/L NaCN
pH of Cyanidization		10.5-11 with $Ca(OH)_2$
Carbon Concentration		10 g/L
Residence Time		24 hours

As reported in the test results, the concentration of arsenic in the residue is 2.1% As. This is considered relatively high, and is probably due to the formation of iron arsenates, which tend to encapsulate gold, and limit its extraction by cyanidization. The formation of iron arsenates may be due to the absence of control on air ratio using a muffle furnace, which is typically essential to slow down the rate of oxidation in the first stage. Fluidization roasting or simulated rotary kiln roasting, controlling air levels by nitrogen doping, was not tested to confirm these assumptions due to budget constraints.

Leaching of the calcine with sodium hydroxide was done to remove arsenic as sodium arsenates. Test results indicate that 33% of the gold is enclosed in arsenic compounds. However, the use of alkaline pressure leaching methods are required, in most cases, to achieve a high extraction of arsenic from calcine. Therefore, the leach results illustrate that arsenic is limiting the recovery of gold.

Optimization studies of gold recovery from the calcine should be performed in order to enhance the economic viability of roasting oxidation. A larger sample of flotation concentrate will be required for fluidized roasting tests. At present, it is assumed that the formation of iron arsenate ($FeAsO_4$) is limiting gold recovery, and that controlling air ratios in a fluidized bed will enable a more efficient volatilization of arsenic.

5.2.4 Pressure Oxidation

For the high arsenic concentrations in the flotation concentrates, acidic-pressure leach conditions were used to extract arsenic. Alkaline conditions, using sodium hydroxide, produce sodium arsenates (Na_2AsO_4), which are very difficult to dispose of. For pressure leaching operations, a complete oxidation of the concentrate and by-products is essential to prevent the formation of elemental sulphur, which may limit the recovery of gold by cyanidization. In addition to elemental sulphur formation, treatment of pyrrohotite and arsenopyrite concentrates, along with the formation of jarosite (which limits silver recovery), has been a limitation for pressure oxidation. In order to quantify the viability of pressure oxidation, the test program conditions were similar to the published Sherritt technology, which has been commercialized on concentrates having similar arsenic contents. The conditions for acidic pressure oxidation are summarized in Table 5.4.

For hydrometallurgical testing, such as pressure oxidation, batch testing is useful for determining the amenability of a particular feedstock. However continuous testing would be required to establish the required level of confidence for a commercial plant. Initially, it was proposed that the testing be performed on a batchwise basis to indicate areas that may be a limitation with the Elmtree concentrate. For the batch test program, the following information could not be determined by small scale testing and would be derived from commercial experience with the Sherritt technology:

- o total consumption of oxygen.
- o heat load on the autoclave, or heat given off by the reaction.

The maximum gold recovery for acid pressure oxidation was 96.1% after six hours of cyanidization. For pressure leaching, a residence time of 120 minutes was required to achieve more than 98% sulphide oxidation. Upon completion of the leach reaction, the residues contained less than 0.2% sulphur, accounting for 94% w/w of feed to the autoclave (6% weight loss by the leach). An increase in the cyanidization leach time was found to have a minimal effect on the recovery of gold.

The autoclave leach liquors and the cyanidization leach residues were tested for arsenic stability. Considerable attention was given to the stability of ferric arsenates ($FeAsO_4$), which may be an environmental concern if the complex is not stable in tailings treatment ponds. Arsenic compounds in the autoclave leach liquors, having an iron to arsenic ratio of 9:1, were found to be stable upon neutralization. However, the residues from cyanidization, having an iron to arsenic ratio of 2.6:1 contained unstable arsenic compounds after neutralization. The test data suggests that the ratio of iron to arsenic is an important design

Table 5.4

Test Conditions for Acidic Pressure Oxidation and Carbon in Leach Gold Recovery

Temperature	190-200°C
Pressure (oxygen)	670 kPa (gauge)
Residence Time	90-120 min.
Liquid/Solid Ratio	5/1 (15% solids w/w)
Electromotive Potential	500-600 mv
Free Acidity	30-50 g/L H_2SO_4
Grind	97%, -400 mesh
Cyanidization Pulp Density	30% solids
Cyanide Concentration	2.0 g/L NaCN
pH of Cyanidization	10 g/L
Residence Time	6-24 hours

criteria for tailings disposal ponds and that the addition of iron (possibly as ferric sulphate) is required for cyanidization residue treatment.

Preliminary amenability tests of the nitric acid catalyzed pressure leach, known as the "Arseno Process", were designed to compare the residence time to conventional pressure oxidation methods. Overheating of the leach resulted in the formation of elemental sulphur and maximum gold recoveries of 89.7%. It was evident from this test that nitrate oxidation requires a leach vessel having adequate cooling capabilities (high heat removal for a short reaction time). The test results also indicated that a lower residence time may be achieved with the Arseno Process. However, the design of a reaction vessel would be required to identify a true residence time and optimum conditions for gold recovery.

From the nitrate catalyzed pressure leaching tests, it is interesting to note that a significant amount of gold reported to the autoclave liquors, as a possible consequence of high temperatures and excessive reaction times. However, the gold content of these liquors would be lost upon solid-liquid separation of the process mixture, downstream of autoclave operations. This loss may be a limitation to highly oxidizing chemical treatment of concentrates and should be quantified with the development of optimum conditions before final conclusions are made.

5.2.5 Bioleach Oxidation

The strain of bacteria commonly used for the oxidation of high arsenic concentrates is known as "Thiobacillus Ferrooxidans", or "TF". As reported in the literature, TF has been shown to be selective for the oxidation of arsenopyrite. However, the selectivity of the oxidant will be dependent on the complexity of the mineralogy. If arsenopyrite is predominantly enclosed by pyrite, the selectivity of TF may not be observed. Standard conditions for bioleach operations to maintain bacteria survival and the solubility of arsenic complexes have been summarized in Table 5.5.

TF was considered as "stock strain" and was obtained from Eimco Process Equipment Company of Salt Lake City. For batch testing purposes, the strain may not ensure an optimum extraction rate. In order to achieve an optimum strain, a continuous operation will be required to run for a significant period. Therefore, batch testing is believed to give an indication of extraction rates, but not necessarily the optimum.

The batch test program was designed to determine the rate of oxidation at standard operating conditions. Optimization of air enrichment, nutrient consumption, solid-liquid ratio, and agitation was not considered. In order to determine the rate of oxidation, the rate of iron extraction was monitored and correlated with gold recoveries by cyanidization of the residue.

For bioleach oxidation, the maximum recovery of gold was 90.4% after 18 days (as shown in Table 10, page 23 of Volume II). A dormant phase occurs at day 15 (as shown in Figure 4, page 22 of Volume II) with a gold extraction of approximately 85 %. From the slope of the reaction rate curve between 6 and 15 days, the steady state extraction rate can be determined. By knowing the maximum gold extraction required, or the maximum iron extraction, a residence time for biological oxidation of the Elmtree concentrate can be determined.

Conditions for the bioleach are important to maintain a maximum extraction rate. If concentrations of iron and/or arsenic reach to high a level, the rate may slow down. However, both iron and arsenic concentrations remained low, averaging 16 g/L total iron and 6 g/L As, at an EMF of 500-645 mv. Upon completion of the reaction, a weight loss of solids was estimated to be about 20%.

In addition to residence time, agitation and cooling requirements are important design variables for bio-oxidation. These variables could not be measured in the test program due to the small scale of the apparatus. Thus, future bioleach piloting should include studies on agitation-air mass transfer and cooling requirements.

Table 5.5

Test Conditions for Bioleach Oxidation and Carbon in Leach Gold Recovery

Temperature	35-38°C
Air Enrichment	1% CO_2 27 mL/min/L pulp
Liquid/Solid Ratio	5/1 (15% solids)
pH	1.5
Nutrients	9 K media
Arsenic Concentration	Less Than 121 g/L As
Iron Concentration	Less Than 35 g/L Fe
EMF	500-650 mv
Grind	97%, -400 mesh
Cyanidization Pulp Density	30%
Cyanide Concentration	2 g/L NaCN
pH of Cyanidization	11.0
Carbon in Concentration	10 g/L
Residence Time	24 hours

The consumption of sodium cyanide was reported to be high during a 24 hour cyanidization of the bioleach residue. The total cyanide and lime consumption was reported to be 7.53 kg/mt and 17.8 kg/mt, respectively, for cyanidization of a completely oxidized residue after 18 days in the process.

For a preliminary assessment of the reagent consumption for the neutralization of bioleach liquors, a lime slurry was added to the final liquor in two stages controlled to a pH 5 and 11 respectively. In the first stage, limestone is generally used to neutralize sulphuric acid and precipitate the bulk iron III, copper, and aluminum. Increasing the pH to 11 in the second stage will precipitate the remaining metals such as iron II, magnesium, manganese, and zinc. Consumption of calcium hydroxide for the first and second stages was reported as 24.8 kg/t liquor and 15.6 kg/t liquor, respectively. It should be noted that the addition of flotation tailings, which may contain a significant amount of calcite ($CaCO_3$), to stage one of the tailings treatment will reduce reagent consumption. The optimization of neutralization conditions was beyond the scope of the present assesment.

5.3 Summary of Test Results

The characterization of the Elmtree ore and concentrates has illustrated the refractive nature of the deposit. Natural surface oxidation was not found to have a dramatic impact on the extraction of gold. However, oxidation of pyrrhotite may have influenced the recovery of gold during preconcentration by flotation. In general, the oxidation and cyanidization properties of the Elmtree ore appear to be typical of other refractory ores. Further detailed assessments of the various plant operations would be required to support a commercial design package, specific to the Elmtree ore.

5.3.1 Flotation

The grade-recovery relationships of the surface ore have been reported to be slightly different than previous testing on core samples. In order to assure that the oxidation rate was based on all refractory gold minerals, a high recovery of gold was required during the test program. For a gold recovery of 93% in flotation, the grade of gold in concentrate is 24.4 g/mt Au. This grade and recovery was considered a base case for the preliminary evaluation of oxidation technologies.

Both the grade and recovery of gold in flotation have an effect on the sizing of an oxidation plant and affect the overall process economics. As illustrated in Figure 5.1, for the production of 75,000 troy oz Au/year, the throughput of an oxidation plant is 12.8 mt/hr of dry solids for a concentrate grade of 24.4 g/mt Au. An increase in flotation grade to 40 g/mt will decrease the oxidation plant throughput to approximately 8 mt/hr, which represents a 37.5% reduction in capacity. However this grade ia achieved at the expense of overall recovery. Assuming a 0.7 power scaling factor for plant costing comparisons, the reduction in capital for an oxidation-cyanidization plant could be expected to be approximately 28% for a 40 g/mt concentrate.

The effect of carbonate minerals in the ore, or elements known as acid consumers, was not determined. Concentrations of calcite ($CaCO_3$) were reported in previous studies as 18 to 20%, which may have an adverse effect on flotation performance, in addition to the consumption of acid during oxidation. However, as pointed out earlier, the presence of acid consumers in the ore may assist in tailings neutralization. Thus, the presence of gangue minerals will have economic implications on operating costs.

5.3.2 Oxidation

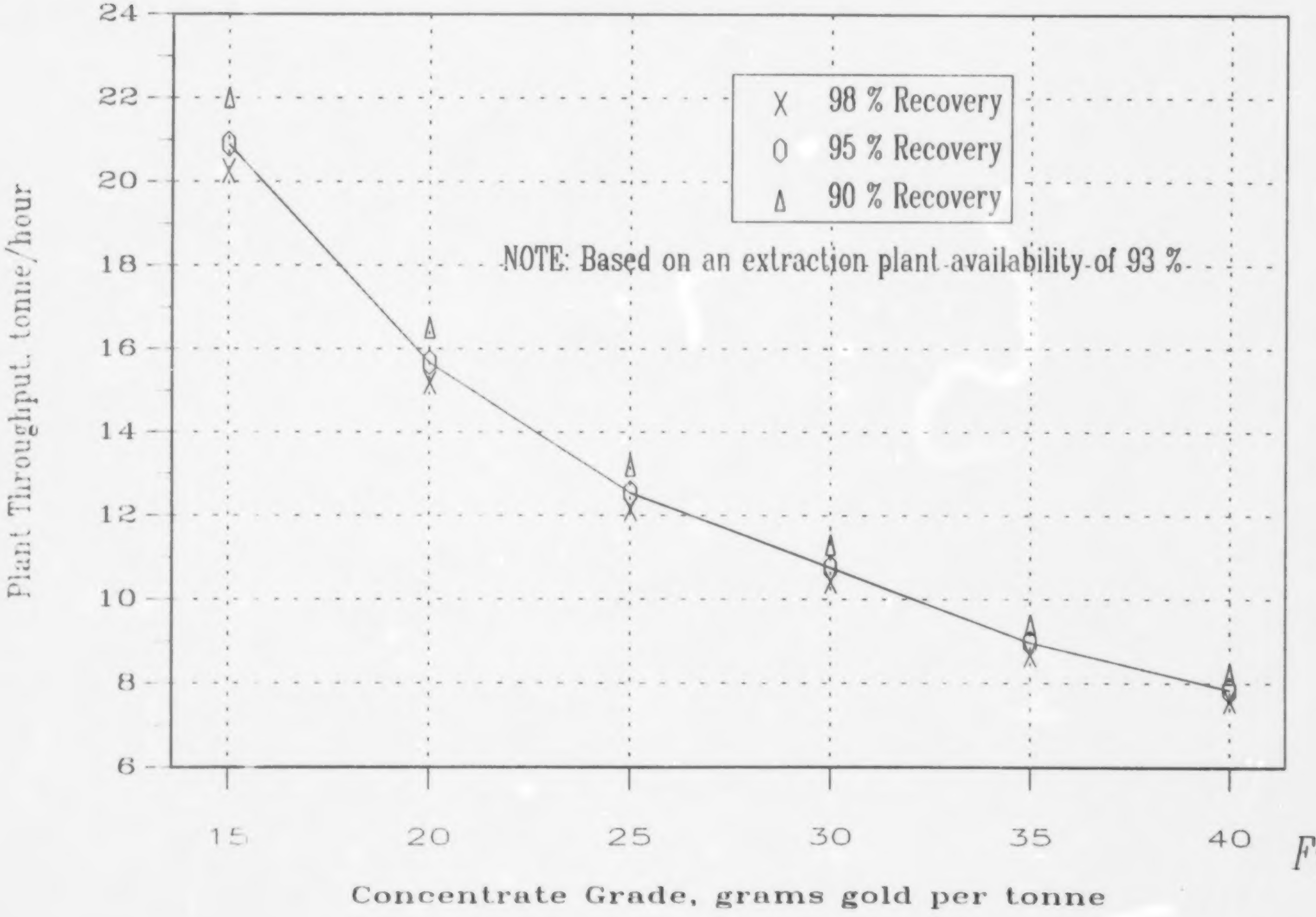
Acidic pressure oxidation technology has demonstrated the ability to achieve high gold recoveries. However, the conditions for oxidation have not been optimized, and the implications of acid recycle have not been quantified. Temperature, for example, may be considered too high for a commercial operation based on equipment metal fatigue. However, the batch test results are considered consistent with commercial plant process conditions, which reduces the technical risk.

The gold recoveries from roasting technology have been low, but may be economically viable. Some commercial fluidized roasters operate at recoveries in the range of 75 to 85%. An explanation for the low gold recoveries from the Elmtree concentrate is not fully understood. However, test data indicates that ferric arsenate may be a problem due to the roasting conditions. Further development will be required to determine ideal temperatures and air ratios for fluidization. If roasting conditions cannot be improved, the use of additional unit operations to chemically oxidize the calcine might be considered.

The recovery of gold by bio-oxidation, in the range of 90 to 95%, is considered standard for this technology. Optimization of the process conditions may give a slightly improved recovery, however, the test results have indicated that a high extraction of iron (approximately 70%) is required to give an acceptable recovery of gold. For some ores and concentrates, only 15 to 25% of iron extraction is required to achieve satisfactory results. Therefore, the necessary time for extraction, in the case of the Elmtree ore, may be considered long by industrial standards. A more accurate assessment by continuous operation of the bioleach circuit would be required to determine a residence time for commercial design purposes.

5.3.3 Cyanidization

The residence time and the solid throughput are related to the scale of operation for cyanidization. In the case of pressure oxidation, the necessary cyanidization was determined to be six hours for preliminary carbon-in-leach assessments. For roaster calcine and bioleach residues, the optimum time for cyanidization was not determined. However, the weight loss of solids using bioleach and roasting technologies was greater than pressure oxidation, thereby decreasing the required size of the cyanidization circuit. An assessment of



Variation in Plant Throughput
Annual Production of 75,000 oz per year

NOTE: Based on an extraction plant availability of 93 %

Figure 5.1

cyanidization residence times, carbon loadings, etc. would be required before the actual size of the cyanidization circuit is determined for comparison.

The consumption of sodium cyanide was found to be high for each oxidation technology. This may be due to the efficiency of washing the oxidation residues. A more detailed appraisal of residue chemistry would identify the types of precipitates and washing requirements. As well, the reduction in cyanidization reaction times will decrease sodium cyanide consumption.

5.3.4 Tailings Treatment

The disposal of arsenic in effluent solutions could present an environmental problem if not properly treated. Various studies have been done on the long term stability of calcium and iron arsenates. The formation of iron arsenates has been determined to be the most stable arsenate. However, the stability of the iron arsenate complex is related to the iron/arsenic ratios. Basic ferric arsenate compounds of molar ratios less than 1 will decompose to ferric hydroxide $Fe(OH)_3$ and soluble ferric arsenate ($FeAsO_4$). As indicated by the test program, oxidation waste liquors may have sufficient concentrations of ferric ions to allow for the formation of stable iron arsenate complexes, but iron in the solid residues may not. Therefore, further development of a waste management system would be required to ensure environmental acceptance of chemical oxidation methods.

The treatment of cyanide that may occur in process effluents was not assessed. It is believed that treatment of these effluents with peroxide or sulphur dioxide technology should be characterized to identify possible limitations with interfering ions. However, for the objectives of the present study, it is assumed that a conventional sulphur dioxide-cyanide removal process would not present any technical difficulties that are unique to oxidation technologies.

6.0 CONCEPTUAL COMMERCIAL PLANT DESIGN AND ECONOMICS

In order to assess the preliminary feasibility of a custom mill, a conceptual equipment scale-up to produce 75,000 troy oz/year from arseno-refractory ores was developed. As indicated in Section 5 of this report, the use of preconcentration by flotation and the extraction of gold by carbon-in-pulp technology have been included in the sizing and cost analysis.

Production costs of various gold operations have been reported in the literature, having a range of 150 to 456 Cdn dollars per troy oz of gold produced. It is also reported that a typical gold operation produces about 85,000 troy oz/year at a production cost of 240 to 290 Cdn dollars per troy oz of gold produced. As a result, although the optimum scale of operation was not determined, the selected production rate of 75,000 troy oz/year would appear to be consistent with present commercial operations.

For a custom mill, the most viable business plan would probably be based on a fully integrated complex for the treatment of ores at the mine site. In addition to total integration, the following development options have been identified for treatment of various refractory gold occurrences having an 18 mile radius surrounding the Caribou Mine of Northern New Brunswick:

- The construction of a greenfield custom mill designed to treat flotation concentrates, produced from surrounding mines and/or imported from off-shore.
- The construction of a fully integrated processing facility at a mine site, having extra capacity in the oxidation and cyanidization circuits to treat flotation concentrate from surrounding mines.

The integrated processing plant, designed to full capacity, has been chosen as a base case for evaluating the order of magnitude capital and operating costs of all three technologies. In addition, production costs for a preconcentrator and a separate custom mill treatment plant have been included.

6.1 Process Flow Description

The process flow drawings for various sections of an integrated and/or greenfield plants are included in Appendix I. At the battery limits of the plant, ore is considered as feedstock or in the case of a custom mill treatment plant, concentrate would be the feedstock. Waste treatment will be required for oxidation waste liquors (from biological or pressure leaching), flotation tailings, and cyanide treatment of solid residue from cyanidization. Process flow diagrams for an oxygen plant (pressure oxidation) and a sulphuric acid plant (roasting) have not been included, since these are outside the defined battery limits.

Preconcentration

Run of mine ore is stored in a silo having a capacity of 4,500 mt. Crushing is assumed to be performed on a seven day a week basis to reduce the feedstock to a particle size of 80% -12.7 mm. Crushed ore is then further ground by a primary rod mill and a secondary rod mill in a closed circuit to 85% -74 microns. A work index of 13.3 kW·hr/st is assumed for the ore, as reported by the Department of Natural Resources and Energy from preliminary grinding studies.

Bulk flotation is accomplished by use of a conventional flotation circuit. A tailings regrind has not been included in the flotation circuit but should be considered in a commercial design. Flotation concentrate is thickened prior to oxidation. For bioleaching, removal of organic flotation reagents by carbon in pulp treatment may be considered if organics are found to be detrimental to the bioleach process. Prior to chemical oxidation, the flotation concentrate is reground to 97% -400 mesh to improve on reaction kinetics and in the case of roasting technology, the calcine is reground to liberate gold.

Pressure Leach Oxidation

In the case of pressure leach oxidation, the flotation concentrate is mixed with acidic recycle solution and preheated prior to entering the pressure autoclave. The slurry entering the autoclave may enter the first chamber at 45%

solids and is diluted to 15% by the addition of water to the autoclave. Lignosol may be used to prevent elemental sulphur buildup on the surface of sulfide particles.

Oxygen is added to the autoclave slurry at elevated pressures to maintain an over pressure of 685 kPa (gauge). The required purity of oxygen is approximately 95-98% and is used instead of air to maintain an autogeneous heat load in leach at 195° to 200°C. Generally, the oxygen would be purchased "over the fence" from a modular cyrogenic oxygen plant.

The slurry is discharged from the autoclave through pressure reducing vessels and is cooled prior to washing.

Bioleaching

The recycle acidic solutions, containing the active bacteria, are mixed with the solids from flotation for oxidation reactions in continuous stirred tank reactors (CSTR's). As a result of oxidation, heat is given off and cooling coils coupled with cooling towers would be required to maintain the slurry at 35°C. The oxidized residue from the CSTR's is washed.

Roasting

A high density slurry of flotation concentrate is fed to the first stage fluidized bed reactor (FBR) at approximately 75-80% solids. In an overflow arrangement, the solids move from the first stage to the second stage FBR and are then quenched in cooling tanks before washing. The majority of the entrained solids in the off-gas are removed by cyclones and an electrostatic precipitator. The off-gas is then cooled to sublime arsenic oxide, which is collected in a bag house.

The acidic off-gas from both roaster units is treated by scrubbing and catalytic conversion of sulphur dioxide to sulphur trioxide. The resulting sulphur trioxide is reacted with water to produce sulphuric acid. Scrubbing efficiencies are in the range of 95 to 98% sulphur capture.

Residue Washing

The oxidation residue is washed to remove soluble forms of iron, arsenic, and sulphate. Washing is generally performed in a counter current decantation circuit (CCD) using two thickeners. In the case of roaster calcine, a single stage washing with regrind of calcine may be effective.

Carbon In Pulp

The washed oxidation residue is treated with cyanide in a cyanide leach circuit. A cyanide leach time of up to 36 hours in six stages is assumed. The slurry from the leach circuit is mixed with coarse activated carbon and slurried in the carbon in pulp tanks.

Activated carbon is transferred from each tank counter current to the flow of the slurry and is separated from the slurry by carbon interstage screens. The carbon, loaded with gold and other cyanogens, is transferred to the strip-electrowinning circuits.

Activated carbon, saturated with volatile metals and carbonates, would be reactivated in a rotary kiln at 600 to 800°C. The entire carbon handling circuit would be designed to remove fine carbon to prevent the loss of adsorbed gold.

Stripping and Electrowin Circuit

Gold is stripped from carbon by a solution of caustic and cyanide under pressure at 150°C for 6 to 8 hours. The strip solution is fed to electrowinning cells that contain steel wool cathodes. Gold is electroplated on the steel wool and refined in an induction furnace.

Tailings Treatment

Limestone and/or flotation tailings is added to waste liquors from oxidation to precipitate heavy metals at a pH of 5. Further precipitation of heavy metals is accomplished in a second step by the addition of lime to maintain a pH of 11.

Cyanide contained in cyanidization residues, is destroyed by the addition of sulphur dioxide, copper sulphate, and lime before discharging to the tailings pond.

After neutralization and cyanide removal, all solids are thickened and, by gravity discharge, are fed to a tailings disposal pond. Water from the tailings pond is reclaimed for process use.

6.2 Process Design Perspective

For each of the three oxidation technologies, preliminary design calculations for the production of 75,000 troy oz/year have been prepared, based upon an integrated plant operation. For each plant, the throughput has been calculated based on an assumed plant availability and recovery of gold as determined in the bench test program. An annual operating period of 351 days and a two week shut down for maintenance were assumed. In addition, the integrated plant throughput was calculated using an ore grade of 5.71 g Au/mt ore. The custom mill facility is derived from these estimates by elimination of the concentrate preparation portion of the integrated plant.

Wherever possible data from the test program has been used as preliminary design criteria. Other design variables have been based on published information and generally accepted engineering practice. In addition, all design calculations are simplified for an order of magnitude sizing of equipment. The preliminary sizing calculations are included in Appendix B through Appendix E.

6.2.1 Acidic Pressure Oxidation

The throughput for the acid pressure leach integrated plant has been calculated at 519,618 mt/year of ore for an overall availability of 88.0% and an overall gold recovery of 89.4%. The pressure leach integrated plant has the highest efficiency of the three oxidation technologies studied, based on availability and recovery, and therefore, the process has the lowest ore throughput requirements.

Operating conditions for pressure oxidation have been based on the lab test results. A reaction time of 120 minutes was required to achieve a gold extraction in excess of 95%. In order to achieve an optimum solubility of reaction products, a solid liquid ratio of 5:1 was selected. The consumption of oxygen and solids settling rate (for CCD thickeners) was based on operating experience of Sherritt Gordon.

Pressure Autoclave Vessel

The pressure autoclave is constructed from a carbon steel liner having titanium alloy components and lined with lead bricks. The sizing calculation was based on a length to diameter ratio of 5.4:1, with the slurry occupying 50% of the volume. On this basis, the vessel size was estimated to be 3.97 m in diameter by 21.45 m in length for a throughput of 13.43 mt/hour of dry solids.

Oxygen Plant

The consumption of oxygen was based on an average of 14% sulphur in the flotation concentrate and a consumption of 2.3 mt O_2/mt sulphur. Sizing of the oxygen plant was therefore based on 100 mt O_2/day at 95 to 98% purity.

6.2.2 Biological Oxidation

Based on standard metallurgical industry practice, the overall availability of the biological oxidation plant was assumed to be approximately 90% although no actual commercial operations have been constructed to date. This assumption was based on the conclusion that the operation of the CSTR's would not be unique and the only outage time would be related to the loss of active biomass. In the event of a major loss of biomass, it may take 7 to 10 days to regenerate the bacteria population. Such an outage would reduce the availability of the plant to 80 to 85% over a year.

The overall efficiency of the process for gold recovery from the feedstock ore based on availability and bench test results was estimated to be approximately 75%. The lower throughput efficiency, compared to pressure oxidation is due to a lower recovery of gold. As a result, the bio-oxidation integrated plant was sized to process 540,533 mt/year of ore.

Residence Time and Bio-reactor Sizing

A maximum rate of extraction of 4.98% Fe/day was determined from the bench scale program (curve shown in Figure 4, page 22 of Volume II). As a result, to achieve a gold recovery of approximately 90%, the estimated residence time would be 13.9 days (333 hours).

Tank sizing was based on conventional fabrication sizes. For a 12.2 m diameter by 12.2 m high tank, a total of 20 tanks would be required for a dry solid throughput of 14.0 mt/hour. The installation of these fully enclosed tanks would include a concrete bottom and a carbon steel siding insulated for outdoor operation.

Due to the lack of commercial experience, the scale up of bioleach CSTR's from batch extraction rate data remains somewhat uncertain. Based on the published literature, most bioleach operations report residence times of less than five days. Recent pilot testing by Gencor Process Research in South Africa accounted for a potential drop in retention time over batch data, due to adaptation of bacteria. In the case reported, it was indicated that the pilot plant was initially sized for 10 days retention time and after two years of researching optimum process variables, the residence time was reduced to four days. Clearly, this demonstrates the need for further continious piloting to establish commercial design specifications. If the residence time were to drop from 333 hours to 133 hours, as experienced by Gencor, the tank volume requirements would decrease by 60%. However, for the present preliminary feasibility assessment, the design was based on the bench scale data.

Agitation and Air Sparging

Engineering design criteria for relationships between slurry agitation and oxygen mass transfer, are an area that is considered proprietary. However, various engineering firms have been involved with pilot programs and have developed expertise in agitation although scale-up has not been proven for commercial operations.

The maximum horsepower requirement for agitation is related to tank size and the amount of agitation achieved by air sparging. The determination of agitation power requirements may be less than standard slurry agitation requirements, therefore, the power rating of 250 Hp per CSTR may be considered conservative. An agitator having 60 to 100 Hp for a 12.2 m tank may be sufficient, but would have to be tested to ensure that agitation does not effect the rate of oxidation.

The consumption of oxygen has been based on lab data and was found to be 0.27 m^3 O_2/m^3 slurry/hour. For a commercial plant, the mass transfer coefficient of oxygen has not been reported in the literature, however oxygen utilization has been measured in pilot programs at approximately 50%.

Heat Load

A number of assumptions have been made on the cooling requirements for bioleach exothermic reactions. In order to maintain biomass regeneration, a temperature of 33 to 37°C must be maintained in the CSTR. The sizing of the heat exchange area was based on a heat input of 920 Kcal/kg solid (reported by Giant Bay Biotech Inc., based on pilot data), and a heat loss for summer and winter operations has been estimated respectively as 10% and 35% of heat input by oxidation. Therefore, the net heat generated by bio-oxidation was estimated to be 11.56 MM Kcal/hour for summer operation and 8.34 MM Kcal/hour for winter operation.

6.2.3 Fluidized Bed Roasting

Fluidized bed roasting has a throughput efficiency of 62.0%, and was the lowest of the three technologies evaluated. This is a result of low availability on the roaster coupled with low gold recoveries. Down time on roasters is generally due to outages in the off-gas handling solids transport systems. As a result, the annual throughput requirements for the roaster (based on an integrated plant) was estimated at 658,688 mt/year ore.

The sizing of the fluidized bed roaster was based on published data from Dorr-Oliver and was factored for a feedrate of 17.02 mt/hour. Temperature, residence time, and air ratios were not determined by the lab test program.

Roaster Sizing

The size of the roaster is dependant on solid throughput, fluidization requirements, and reaction residence time. Sizing parameters are based on testing done by Dorr Oliver using arseno-refractory concentrates from the Lancefield Mine, Windurra, W.A. (which are similar to Elmtree concentrates). A two stage system is employed, having different design parameters. As a result, the diameter of the primary roaster was estimated to be 5.5 m with the secondary roaster at 3.6 m.

Off-Gas Treatment

The total volume of off-gas from both roasters was estimated to be 11,620 N m^3/hr, calculated from fluidization requirements. In order to remove arsenic in the flue gas, it first sublimed by cooling the gas stream to form arsenic trioxide, which is then collected in the bag house. Cooling the off-gas may be accomplished by a waste heat recovery boiler, water addition, and/or by quenching the gas stream with ambient air. As indicated by calculations of cooling air based on thermodynamics, the introduction of cooling air will reduce the off-gas concentration in the range of 3.4 to 5.2% w/w SO_2 based on 11 to 17% sulphur in the concentrate.

The capital and operating costs of a 98% efficient SO_2 scrubber is inversely proportional to the concentration of sulphur dioxide in the gas. Neutralization of the gas with lime scrubbing would require a significant amount of lime on a daily basis, and would not likely be economical. Instead of a throw-away plan for off-gas scrubbing, the production of sulphuric acid is considered more viable for the process.

For the economic production of sulphuric acid from off-gas, the concentration of SO_2 should be in the range of 8-12% w/w. Therefore, an alternative to gas quenching should be considered for cooling as a part of a detailed engineering design study. For present purposes, it was assumed that cooling the gas to achieve sublimation of arsenic and to maintain a gas concentration of SO_2 at 8%, can be accomplished by a combination of air and water quenching.

Heat Load

Auto thermal heat calculations were not carried out based on slurry feed to the reactor. However, it was assumed that the reactor will run autogenously at 17.4% sulphur in the feed using a 10% w/w water content in the feed based on commercial industrial experience. The net heat load on the reactor will be approximately 6.4 MM Kcal/hour, and if the sulphur content of feed decreases to approximately 11%, the reactor will require an additional 2.5 MM Kcal/hour auxilliary heat source.

6.2.4 Primary Upgrading

Power requirements for crushing have been factored from similar operations. The net power requirements for primary, secondary, and tertiary crushing were estimated to be 2.8 kW·hr/mt.

Power requirements for primary and secondary grinding have been calculated based a work index of 13.3 kW·hr/st. The net power requirement for grinding was 16.9 kW·hr/mt. The diameters of the required rod and ball mills have been calculated to be 2.7 m and 3.7 m in diameter, respectively, for both the acidic pressure and bio-oxidation processes. However, slightly larger mills will be required for roasting technology due to higher throughput requirements.

Flotation cells were sized based on commercial cell kinetics for the flotation of arsenopyrite and pyrite ores. Total cell volumes of 69 to 88 m^3 were estimated for rougher, scavenger, and cleaner unit operations. Details of the sizing calculations are summarized in Appendix C.

6.2.5 Carbon in Pulp (CIP)

Preliminary design for cyanidization leaching and carbon movement/treatment systems were based on commercial processing variables. For each oxidation technology, the throughput of solids may vary due to variations in the design tonnage requirements for gold recovery and the variation in weight loss of solids in oxidation. Therefore, to reduce the number of calculations, an average feedrate of 11.8 mt/hr was used as a design basis. The estimated sizing error for bioleach and pressure oxidation is ±6%. For roasting, the CIP circuit was assumed to be 21% larger than average. Details of the sizing calculations are summarized in Appendix D.

6.2.6 Tailings Treatment

Sizing of the neutralization circuit was based on acidic pressure oxidation installations. The destruction of cyanide using SO_2 methods was based on engineering design parameters reported by Inco.

6.3 Process Economics

Order of magnitude capital and operating costs were developed for a total integrated plant based on the preliminary conceptual design. The cost estimated for a greenfield concentrator and a separate custom chemical treatment plant was derived from the integrated plant costs. These cost estimates were prepared for the purpose of evaluating and comparing the preliminary economic feasibility of various technological options for a custom mill. In addition, the assessment of the process economics will provide a basis for development of the custom mill concept.

6.3.1 Capital Cost Estimates

The process parameters presented in Appendices B to E were used to derive approximate equipment sizing for all major components included on the conceptual flowsheets. The estimated purchase cost for major items was based on budget pricing from major equipment suppliers. The balance of the equipment was costed by applying generally accepted industry estimating guidelines, adjusted for the projected size and duty. Conversion to a third quarter 1988 basis was calculated by factoring these estimates against the published Marshall & Swift Equipment Cost Index or the Chemical Engineering Plant Cost Index. Details of the capital costing are summarized in Appendix F. Budget Quotes from equipment suppliers were included for the following major equipment:

- o pressure autoclave, refractory lining, agitators, heat exchangers, and flash tanks for pressure oxidation.

- o fluidized bed reactors, burners, blowers, cyclones, refractory, precipitator, ducting, controls, feeders, quench tank, and bag house for fluidized bed roasting.

- o tanks, corrosion coating, insulation, and agitators for biological oxidation.

In addition, the following components of the total capital cost were factored from the estimated purchase of equipment (POE) cost:

- o Instrumentation @ 10 to 15% of POE based on the relative complexity of the process control requirements.

- o Piping @ 10 to 25% of POE based on the number of chemical processing unit operations.

- o Electrical @ 15 to 22% of POE based on relative complexity of motor control requirements.

- o Services and utilities @ 10% of POE.

- o Engineering design and supervision of construction @ 19 to 25% of POE of total direct cost.

- o Construction indirects @ 3% of POE, commissioning and spare parts @ 2% of POE and a contingency @ 10% of POE.

- o Working capital @ 10% of total fixed cost.

Building and land costs are based on a tentative equipment layout and plant battery limit requirements.

On this basis, the overall capital cost to design and construct a fully integrated gold processing plant was then estimated for each technology. Table 6.1 provides a breakdown of the major components included in this cost. In addition, the total capital cost for a greenfield concentrator and a separate custom chemical treatment plant is summarized in Table 6.2 and 6.3, respectively. The total capital investment does not include the following:

- o Process development cost (either licensing or research and development expense).
- o Electricity generation and distribution outside the plant site.
- o Infrastructure for transportation, waste disposal, and other non-process related items.

6.3.2 Operating and Maintenance Cost Estimates

For the purpose of deriving an estimate of production costs, exclusive of capital charges, a series of key assumptions were made:

- o Operation of the plant is 7 days per week, 24 hours per day.
- o For each case, staffing levels have been derived for a stand-alone plant. The average employee wage was determined to be approximately $33,240 per year and an allowance was made for 32% payroll burdens and 10% overtime expense. Total staffing for an integrated plant was estimated at 98 persons with a reduction to 70 for a greenfield chemical treatment plant.
- o The cost of plant consumables was based on the following:
    - lime @ $94/mt
    - limestone @ $17/mt
    - sodium cyanide @ $1.98/kg
    - oxygen @ $67.25/mt O_2
    - flocculent @ $3,130/mt
    - liquid SO_2 @ $250/mt

Table 6.1

Total Capital Investment for Integrated Plant

75,000 Troy Ounces per year Gold

	Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
		(1988 $CDN X 1000)	
Purchase of equipment	19,379	14,922	18,524
Installation cost	10,125	8,960	12,596
Instrumentation	2,907	1,492	2,779
Piping	3,876	3,730	1,852
Electrical	3,876	3,283	2,779
Plant Services/Utilities	1,938	1,492	1,852
Building/Yard	1,150	1,475	1,150
Land	120	120	24
Total Direct Cost	43,371	35,474	41,546
Engineering & Supervision (20%)	8,674	7,095	8,309
Construction Indirects (3%)	1,301	1,064	1,246
Commissioning, Spare Parts (2%)	867	709	831
Contingency (10%)	4,337	3,547	4,155
Total Indirect Cost	15,179	12,415	14,541
Total Fixed Cost	58,550	47,889	56,087
Working Capital (10%)	5,855	4,789	5,609
Total Fixed Capital Investment	64,405	52,678	61,696

Level of Confidence (-5, + 30%)

CE Index: 348.3

() Percent of Total Direct Cost

TABLE 6.2

Total Capital Investment For Primary Upgrading

(Greenfield Concentrator)

	Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
	(1988 $CDN X 1000)		
Purchase of equipment*	4,413	4,535	5,141
Installation cost	2,113	2,171	2,468
Instrumentation (10%)	441	453	514
Piping (20%)	882	907	1,028
Electrical (20%)	882	907	1,028
Plant Services/Utilities (10%)	441	453	514
Building/Yard*	460	460	460
Land**	48	48	48
Total Direct Cost	9,680	10,064	11,201
Engineering & Supervision [20%]	1,936	2,013	2,240
Construction Indirects [3%]	291	302	336
Commissioning, Spare Parts [2%]	194	201	224
Contingency [15%]	1,452	1,510	1,680
Total Indirect Cost	3,873	4,026	4,480
Total Fixed Cost	13,553	14,090	15,681
Working Capital, 10% Fixed Cost	1,355	1,409	1,568
Total Fixed Capital Investment	14,908	15,499	17,249

Level of Confidence (-5, + 30%)

CE Index: 348.3

* requirements for crushing, grinding, flotation, and concentrate thickening are included.(concentrate packaging and tailing treatment not included).

** factored as 40% of building, yard, and land requirements for a fully integrated plant.

() percentage of purchase equipment cost.
[] percentage of total direct cost.

TABLE 6.3

Total Capital Investment For Custom Mill

(Greenfield Treatment Plant)

(1988 $CDN X 1000)

	Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
Purchase of equipment (1)	14,966	10,687	13,383
Installation cost	8,012	6,789	10,118
Instrumentation*	2,245	1,069	2,007
Piping*	2,993	2,672	1,338
Electrical*	2,993	2,351	2,007
Plant Services/Utilities*	1,497	1,069	1,338
Building/Yard*	690	885	690
Land	72	72	14
Total Direct Cost	33,468	25,594	30,895
Engineering & Supervision [20%]	6,694	5,119	6,179
Construction Indirects [3%]	1,004	768	927
Commissioning, Spare Parts [2%]	669	512	618
Contingency [15%]	5,020	3,839	4,634
Total Indirect Cost	13,387	10,238	12,358
Total Fixed Cost	46,855	35,832	43,253
Working Capital (10%)	4,686	3,583	4,325
Total Fixed Capital Investment	51,541	39,415	47,577

* Based on percentage of purchase of equipment cost as outlined in Appendix F.3.0.

** Factored as 60% od building, yard and land requirements for a fully integrated plant.

(1) Equipment does not include concentrate feedstock handling systems.

[] Percentage of total direct cost.

- electricity @ $0.0425/kW·hr

o Purchase of ore (as run of mine ore) is based on an O&M cost of $CDN 20/mt ore and a capital cost of $CDN 30 million for a 1,000 to 1,500 mt/day operation.

o O&M cost for a sulphuric acid plant is estimated to be $CDN 40/mt sulphuric acid.

o Maintenance cost is factored as 3% of direct cost and operating supplies is 20% of labor cost.

o Insurance is based on $0.35/$100 of direct cost.

Tables 6.4, 6.5, and 6.6 present a summary of the total direct operating costs for each technology. Based on the assumptions used, the major operating costs for the treatment of ore/concentrate are associated with labor requirements and the purchase of feedstock and reagents.

6.3.3 Preliminary Economics

Given the grade of the Elmtree ore together with the current depressed price of gold, none of the processing options evaluated yields an economic investment opportunity at the present time. However, it has been possible to perform a comparative analysis of the three options studied, in order to identify a preferred development direction.

The comparison of oxidation technologies has been performed in detail utilizing *Kashpak*, an economic and financial assessment model developed by The SNC Group for the Department of Natural Resources and Energy. In addition, a simplified levelized cost comparison is also presented.

TABLE 6.4

Operating and Maintenance Costs For Integrated Plant

75,000 Troy Ounces Per Year Gold

		Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
			(Annual Cost 1988 $CDN X 1000)	
A.	Labor Costs (98)	5,084	5,084	5,084
B.	Reagent Costs	7,444	5,001	1,852
C.	Ore Purchase	15,521	16,146	19,676
D.	Acid Plant	-	-	1,970
E.	Electricity	910	1,896	1,084
F.	Operating/Maintenance Supply & Insurance	2,470	2,205	2,408
G.	Refining Charge	719	719	719
	Total	32,148	31,051	32,793
Cost Per mt Ore		61.868	57.467	49.796
Relative		1.24	1.15	1.0

NOTE: Taxation, Depreciation Not Included

Plant Maximum Service Factor = 8000 hours/year

Level of Confidence (-5%, +30%)

TABLE 6.5

Operating and Maintenance Costs For Primary Upgrading

(Greenfield Concentrator)

		Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
		(Annual Cost 1988 $CDN X 1000)		
A.	Labor Costs (32)*	1,661	1,661	1,661
B.	Reagent Costs	570	570	570
C.	Ore Purchase (from mine)	15,521	16,146	19,676
D.	Electricity	483	502	608
E.	O&M Supplies/Insurance**	656	669	705
F.	Transportation of Concentrate***	662	688	838
	Total	19,553	20,236	24,058
Cost Per mt Ore		37.63	37.45	36.52

NOTE: Taxation, Depreciation Not Included

Level of Confidence (-5%, +30%)

* Labor included overtime, burden/fringe benefits (average annual salary is $51,900)

** Guidelines for O&M are stated in Appendix G.7.0

*** Based on a haul rate of $0.117/km.mt (50 km haul distance)

TABLE 6.6

Operating and Maintenance Costs For Custom Mill

(Greenfield Treatment Plant)

		Acidic Pressure	*Bio-Leach*	*Fluidized Bed*
			(Annual Cost 1988 $CDN X 1000)	
A.	Labor Costs* (70)	3,633	3,633	3,633
B.	Reagent Costs	6,874	4,431	1,289
C.	Acid Plant (O&M)	-	-	1,970
D.	Electricity	427	1,227	476
E.	O&M Supplies/Insurance**	1,848	1,585	1,761
F.	Refining Charge	719	719	719
G.	Cost of Concentrate (1)	22,102	22,887	27,008
	Total	35,603	34,482	36,856
Cost Per mt Concentrate		314.82	293.08	257.23

* Labor included overtime, burden/fringe benefits (average annual salary is $51,900)

** Guidelines for O&M are stated in Appendix G.7.0

(1) Price of concentrate based on capital and operating costs of primary upgrading for a 15% ROI for 15 years.

Detailed financial results from *Kashpak* for six evaluation cases are presented in Appendix H. All cases were evaluated in constant 1988 Canadian dollars, assuming 100 % equity and a fifteen year project life. In view of the negative cashflows for all of the Custom Mill cases, it was considered superfluous to introduce considerations relating to interest during construction and debt leveraging. As a result, all capital expenditures are assumed to occur in the first year of production. Capital and operating costs have been included as per the figures reported earlier in this section. Revenues have been credited using a gold value of $ 480 CDN per troy ounce. In addition, the fluidized bed roaster case has been credited with revenue from sulfuric acid sales at $ 50 per tonne of acid.

From the projected capital expenditure and annual net revenue, the *Kashpak* program was then used to calculate the internal rate of return for each development option. In addition, the sensitivity of the project to increases or decreases in capital and operating costs, as well as revenue, were evaluated over a range of +/- 30 %. Finally, for purposes of comparison, each technology option was evaluated as both a custom mill without primary upgrading and as a fully integrated stand alone entity to determine the degree of economic improvement which could be realized by eliminating a separate intermediate processing plant. It should be noted however, that the only savings credited in this evaluation related to the reduced capital cost which would be possible by eliminating a separate entity. In actual fact, it must be realized that the primary upgrader will also require a profit against the required capital investment which would have to be paid by the custom mill operator.

The results of the *Kashpak* analysis for the base cases for each option are summarized below:

Internal Rates of Return

Technology	*Custom Mill* %	*Integrated Plant* %
Bio-Oxidation	-6.88	5.41
Acidic Pressure Oxidation	-21.1	-1.49
Fluidized Bed Roaster	-8.38	5.01

In terms of economic sensitivities, it has been found that all options are relatively insensitive to capital costs due to the fact that the projected annual operating and maintenance costs represent over 50 % of the total capital investment in all cases. In addition, net revenue, calculated as total revenue less total O&M expenses and depreciation is negative. As a result, a 30 % decrease in the capital cost of a Biological Oxidation custom mill, would result in the rate of return improving from -6.88 % to -3.08 %. However, a 30 % decrease in operating costs or alternatively, a similar increase in the price of gold, would improve the rate of return on the project to +40 to +45 %. For the Fluidized Bed Roaster and Acidic Pressure Leach options, the sensitivities are equally dramatic although the magnitude of the improvement is somewhat less. Clearly, should the price of gold reach $ 600 per troy ounce, the economics of all options would improve, although the Biological Leach technology would remain the most attractive.

It is interesting to note that while the rate of return for an integrated plant is somewhat improved over the custom mill at present gold prices, the sensitivity to increases or decreases in revenue and O&M costs is reduced. This result is due to the higher proportion of capital for such a plant.

The full details of the six cases studied, including the sensitivity calculations, are included in Appendix H.

In order to assist in the understanding of the factors affecting the comparison of the three oxidation technologies, a simplified levelized cost comparison has been prepared. Figure 6.1 presents the results of the custom mill comparison while 6.2 presents a comparison based on an integrated plant including primary upgrading. In preparing these charts, the capital cost of each case was converted to an annual sinking fund payment based on a 15 year amortization period at 6 % annual interest. The interest amount reflects an assumed inflation level, while the operating and maintenance costs have been expressed in 1988 Canadian dollars.

These results clearly indicate that the Biological Leach option results in lower annual costs for both the custom mill and the integrated plant. This is partially due to the lower annual capital payment. In the case of Fluidized Bed Roasting, the major difference relates primarily to the improved gold recovery from feedstock ores or concentrates. This penalty relates to the results of the pilot test program. It should be noted however, that a revenue of 2.46 million dollars per annum has been credited to the Fluidized Bed case to account for the sale of by-product sulfuric acid at $ 50 per tonne. If a market were not available for the acid, the cost of neutralizing and disposing of it would place this option well behind the Acidic Pressure Leach option.

7.0 DEVELOPMENT CONSIDERATIONS

The commercial feasibility of establishing a custom mill for refractory gold processing in Northern New Brunswick, is dependent on technical and economic factors. In this regard, the present assessment has identified the limitations of three technologies which may be applied to such a project. However, the current depressed price of gold coupled with the low grade of the Elmtree deposit evaluated in this study, would appear to render any development uneconomic at this time.

The evaluation completed to date has yielded comparative results which may be of assistance in planning further development initiatives. Clearly, aside from a sharp increase in gold prices, a successful project will require substantial improvement in the operating and capital costs incurred per tonne of feedstock processed. The following subsections identify areas which may warrant further consideration in an effort to optimize the choice of technology for a successful commercial venture.

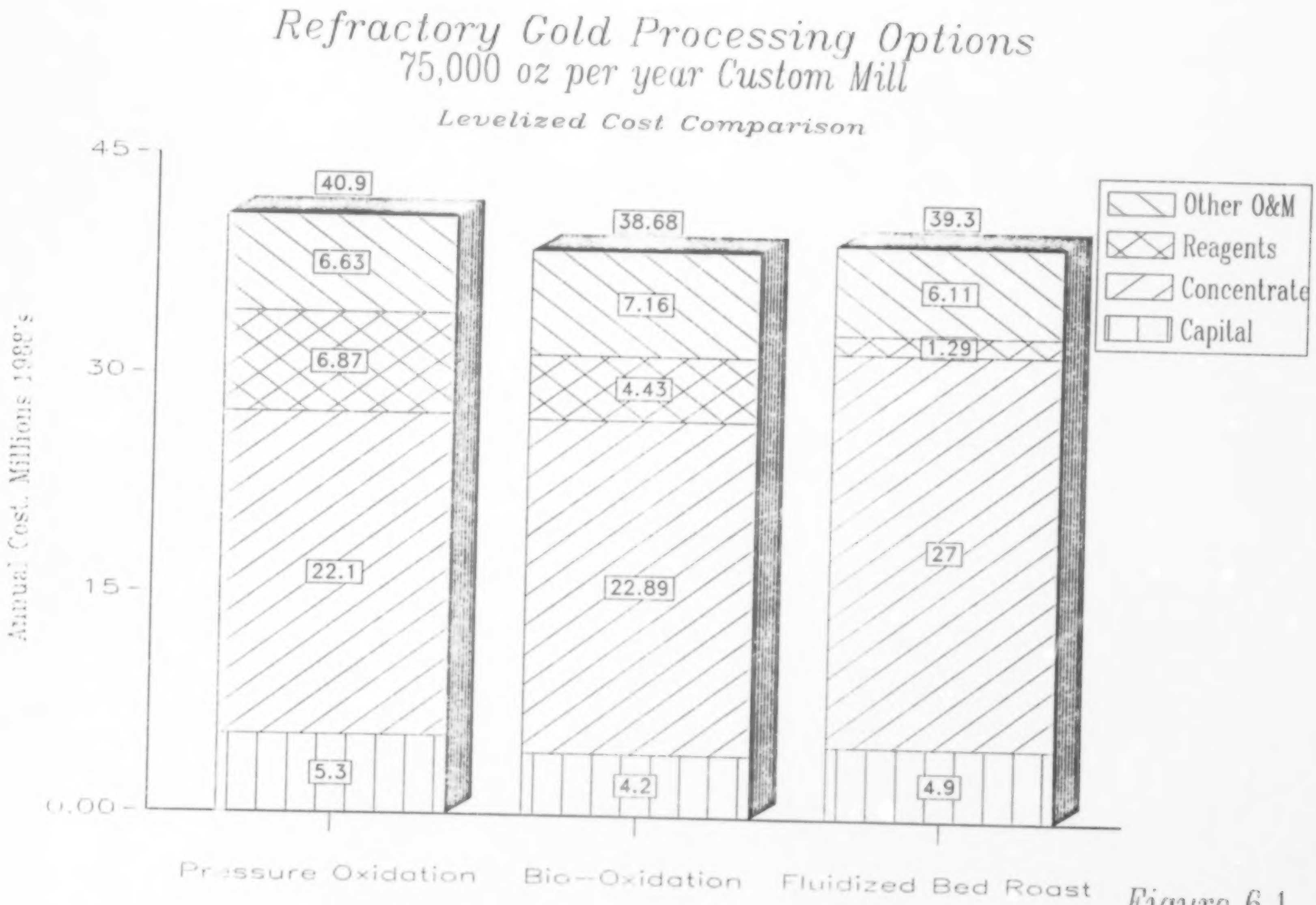


Figure 6.1

NOTE: O&M cost for Fluidized Bed Option is reduced by $2.46 million to account for revenue of $ 50 per tonne of acid produced

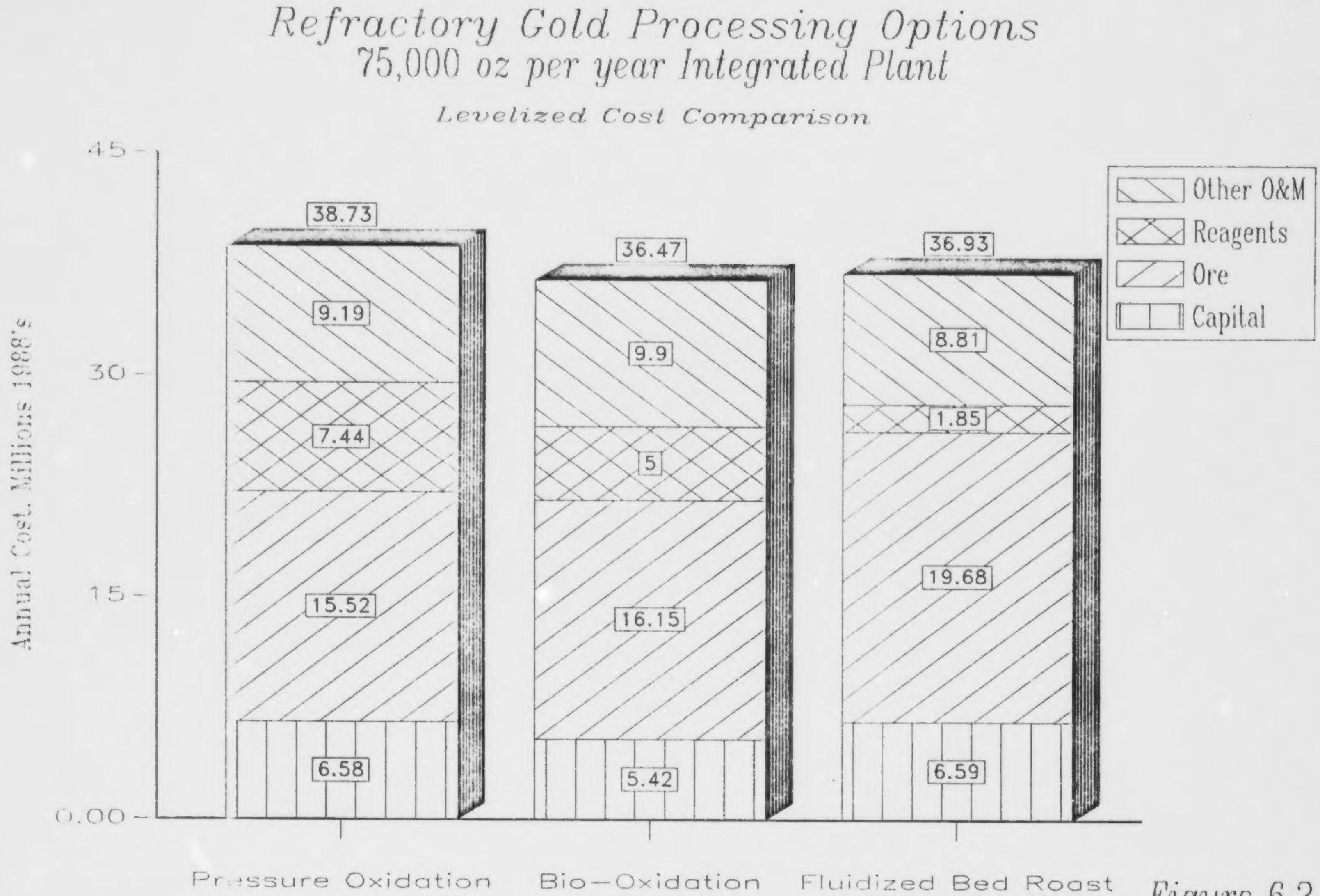


Figure 6.2

NOTE: O&M cost for Fluidized Bed Option is reduced by $2.46 million to account for revenue of $ 50 per tonne of acid produced

7.1 Technical

Based on the laboratory investigations and the preliminary engineering assessment completed during the present study, the most critical technical parameters which must be optimized prior to a final feasibility assessment include not only the selection of the best technology but also the choice of the optimum feedstock.

In terms of the efficiency and technical risk of specific oxidation technologies, the Fluidized Bed Roaster and Acidic Pressure Leach concepts carry a higher degree of confidence than does the Bio-Leach approach. However, the roaster did not prove as efficient in gold recovery, with the result that operating costs were substantially increased due to the higher throughput required for a fixed production level of 75,000 ounces of gold per year. In addition, the high capital cost of Sherritt's pressure leach process makes this approach to oxidation the least attractive. Both of these technologies could benefit from further piloting although the likelihood of achieving significant capital and operating cost reductions are considered minimal. In the case of roasting, continuous testing in a bench scale fluidized bed would ascertain whether control of air ratios could improve the performance of the process.

The Biological leach approach however, does hold the promise of substantial reductions in both capital and operating costs, should the overall residence time be reduced in continuous steady state operations. For this reason, it would be most beneficial to consider further piloting of this technology to more precisely determine not only this critical parameter but also to evaluate heat load and stability of the process.

With regard to emerging processes, it would appear that the potential for a substantial improvement in reaction rate and corresponding throughput increase may be achieved by the Arseno technology. Limited testing conducted by Lakefield Research during the course of this program, confirmed vigorous reaction conditions, although overall yield was not as high as desired. Given the potential to substantially reduce at least the capital cost of a custom mill, further evaluation of this concept may have more potential to yield a commercial project than either the Roasting or Sherritt processes.

In addition to the choice of oxidation technology, it is considered critical to ensure that the feedstock selected strikes the optimum balance between processing cost and grade achieved. In this regard, further study would appear to be warranted to determine the potential for overall economic improvement by optimizing the design of the primary upgrading circuit. These investigations should consider both multi-stage flotation as well as gravity separation circuits. Although the primary processing step is currently envisaged to be outside the scope of a custom mill, it will be critical to the latter in terms of setting feedstock specifications.

Perhaps the most critical technical aspect of the custom mill concept which is not addressed by the present study, relates to the effect of feedstock variability on the stability and performance of the oxidation process. This is considered to be of critical importance to all technical options, however, the biological leach concept may be most susceptible to adverse impacts in this regard. For this reason, it is considered important to consider this variable in future evaluations of refractory gold processing options, particularly for a custom mill concept.

During the laboratory program, it was noted that some concern may exist as to the environmental acceptability of the iron arsenic complexes associated with neutralization effluents arising from all but the roaster process. As the commercial viability of a specific process will be dependent on the ability to meet environmental standards for all effluents, it will be important to identify the precise conditions which will ensure a stable waste material.

7.2 Economics

Between 50 and 60 % of the equivalent annual costs associated with a refractory gold processing facility, are expected to be related to feedstock acquisition. As a result, economics would dictate that every effort be made to ensure that this aspect of the project be carefully optimized. Several factors will have to be balanced in this process including, grade of ore and degree of primary upgrading. The former is a function of the deposits themselves and cannot be controlled, while the latter must be resolved through a cost benefit study. This type of analysis will be more difficult for a custom mill venture where perhaps several external interests will exercise the ultimate control over development directions. However, for a custom mill to be viable, it will be essential to have a coordinated approach to this aspect of the project. In fact, it may be that feedstock costs will have to be more a function of production cost plus agreed profit margin than it is of price of gold, in order for a custom mill to become attractive.

Clearly, another important economic factor, relates to the economies of scale. While the 75,000 ounce per year custom mill considered in the present study, is typical of the average size of present operations, it is perhaps somewhat small for a new venture. The optimum scale for a plant located in Northern New Brunswick, will depend on the availability of reserves and an assessment of the risks and benefits associated with different production capacities.

Finally, the performance risk associated with a specific technology must be included in the evaluation of its commercial merits. It may be best to accomplish this by assigning discount factors to the revenue stream, in accordance with the perceived degree of production risk. For example, the factor assigned to Roasting technology might be 1.0, whereas an emerging technology such as the Arseno process may be penalized with a 0.9 multiplier in recognition of the unproven nature of the technology and the scale up risk.

8.0 CONCLUSIONS & RECOMMENDATIONS

The present study has not been successful in clearly identifying a single technology which is significantly more attractive for the development of a custom mill for refractory gold processing in Northern New Brunswick. However, the following major findings have been developed:

o The currently depressed price of gold, coupled with the low grade of the ore evaluated, makes the development of a plant uneconomic at this time. It is believed that, if nothing else were to change, the price of gold would have to reach approximately $ 600 CDN per troy ounce for the concept to become commercially feasible.

o The single largest component of the cost of constructing and operating a custom refractory gold processing mill, is expected to be associated with the acquisition of a suitable feedstock. Based on the grade of the Elmtree deposit, this cost represents between 50 and 60 % of the total annual costs.

o Of the three oxidation technologies evaluated, the Biological Leach option shows the best overall economics based on laboratory performance results and preliminary engineering scale up. In addition, this technology appears to offer the best hope for significant capital and operating cost reductions through further steady state piloting.

o The emerging Arseno process appears to offer a potential for reduced capital costs and perhaps some operating cost improvement. However, considerable piloting and engineering study will be necessary to achieve an acceptable level of technical risk.

o The economic viability of all processes evaluated are very sensitive to increases or decreases in revenue or operating expenses. Sensitivity to the capital cost of the oxidation plant is a relatively minor factor.

o A preliminary comparison of the custom mill concept with a fully integrated processing facility, indicates that the latter is slightly more attractive. This results primarily from the economies of consolidating operations under one roof, and may be more a function of the nature of feedstock supply contracts for a custom mill. If the cost of feedstock is more a function of production cost plus fixed profit margins than it is of the gold price, the economics of the custom mill may improve with a higher grade concentrate.

The major recommendations arising from the present assessment can be summarized as follows:

o Prior to conducting further piloting tests on any technology, the variability and reserves of refractory deposits in Northern New Brunswick should be compiled and quantified.

o If warranted by reserves, further steady state piloting evaluations should be conducted on the Biological Leach option in order to more precisely establish reaction parameters such as residence time, agitation requirements, heat load and stability.

o If warranted by reserves, a preliminary bench scale assessment of the Arseno process should be conducted to establish the performance of this technology and identify the potential savings which could accrue through commercialization.

o Again, depending on the extent and grade of refractory reserves, some preliminary effort should be dedicated to the identification of the economically optimum grade versus recovery in primary upgrading.

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APPENDIX A

The Geology, Mineralogy and Bulk Sampling
of the
Elmtree Gold Deposit

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1.0 INTRODUCTION

Corona Corporation's Elmtree gold deposit is located approximately 10 kilometres west of Petit Rocher in northeastern New Brunswick. The property is accessed via the Villages of Laplant, Alcida and Dauversière and straddles the South Branch of the Elmtree River. The discovery of the deposit was made in the fall of 1984 by Lacana Mining Corporation which has since been taken over by Corona. Exploration interest in the area stemmed from publication by the New Brunswick Department of Natural Resources of anomalous regional stream sediment analyses results within favourable geological environments. Prospecting in the area of anomalous arsenic and antimony sites in the South Branch of the Elmtree River led to the discovery of semi-massive sulphide-quartz veins carrying gold and silver values ranging up to 17.5 grams/tonne and 555.0 grams/tonne, respectively. Additional prospecting and staking of the so-called Discovery Zone subsequently led to the discovery of the West Gabbro Zone which hosts the Elmtree gold deposit.

A significant amount of exploration work has been carried out on the property. The work includes ground geophysical and geochemical surveys, geological mapping and 10,700 metres of diamond drilling. The bulk of the effort has been aimed at the West Gabbro Zone.

2.0 GEOLOGY

2.1 Regional Setting

Geological studies undertaken by D. Hoy (1988) on and around Corona's mineral holdings provide a detailed description of the geological setting of the Elmtree deposit.

The property is underlain by Ordovician to Silurian, metasedimentary rocks that are intruded by mafic and felsic sills of probable Devonian age.

The Elmtree Group, of Ordovician age, underlies the northern part of the claim group and consists of a sequence of graphitic argillites, phyllites, slates and greywackes. Locally, there are interbeds of coarse grit and pebble conglomerate. Skarn has developed within the contact metamorphic aureoles of intrusive rocks.

The Elmtree Group is in fault contact with a thick sequence of calcareous metasediments belonging to the Silurian, Chaleurs Group. The metasediments consist of greenish-grey siltstone, argillite, sandstone, pebble to cobble conglomerate and minor limestone. The Chaleurs Group generally underlies the southern part of the property.

Locally, intrusive rocks cut the Elmtree and Chaleurs Groups. Stocks, sills and dykes are found consisting of diabase, gabbro, granite, feldspar porphyry and felsite.

The Elmtree Fault is a broad zone of intense shearing, fracturing and deformation which separates the graphitic argillites of the Elmtree Group to the north from the calcareous siltstones of the Chaleurs Group to the south. Virtually all mineralization discovered on the property to date is intimately associated with the fault zone or with parallel and en echelon splays from it. The gabbroic sill which comprises the West Gabbro Zone is thought to have been injected along this zone of structural weakness.

2.2 Deposit Geology

The Elmtree Deposit is located within that portion of the property known as the West Gabbro Zone. The Zone consists of a hydrothermally altered and strongly deformed gabbroic sill, bounded by graphitic argillites and hornfels of the Elmtree Group. The sill has been intermittently traced along a strike length of 460 metres and ranges in width from 3 to 40 metres. The sill strikes approximately 60° and dips are near vertical at shallow depths but assume a more northerly component at depth.

The sill is characterized by fine to medium grained ophitic margins, gradational into a medium to very coarse grained core. Compositionally, the margins are characterized by equigranular gabbro, whereas the core is gabbroic anorthosite to anorthosite and is auriferous; the core of the sill within the West Gabbro Zone forms the Elmtree gold deposit.

Ductile and brittle deformation is locally very intense within the sill. Petrographic analysis indicates that original igneous textures are not well preserved and that the original mineralogy of the gabbros consisted of calcic palgioclase and at least one mafic mineral.

2.3 Mineralogy

The alteration of the gabbros consists of saussuritization of plagioclase, associated with chlorizitization of ferromagnesium minerals. Within highly deformed gabbro, a microcrystalline alteration assemblage of carbonate, quartz, chlorite, albite and sericite forms a matrix interstial to plagioclase and mafic mineral remnants. This alteration assemblage is also commonly seen in several generations of veining.

Higher gold values are associated with sulphide-rich zones cutting the coarse grained gabbro. The zones of highest sulphide mineral content, containing 10 to 25% total sulphide mineralization are proximal to the margins of quartz veins in zones of silicification.

Polished thin section microscopy and microprobe analysis was performed by Miller (1988) on thirty-four samples taken from drill core from various depths in the West Gabbro Zone. These analysis were used to define the mineralogy of the Zone, including the auriferous anorthositic gabbro.

2.3.1 Host Rocks

The mineral assemblage identified in the host rock consists of quartz, feldspar (plagioclase and orthoclase), mica (biotite, sericite), chlorite, calcite and dolomite. This assemblage is characteristic of greenschist metamorphism. These metagabbroic rocks, which hosts the majority of the sulphides and contains the significant gold values, ranges from fine grained to coarse grained. The coarse metagabbro, contains grains up to 2 centimetres, and is generally altered and brecciated. Carbonatization occurs throughout the host rocks and calcite and dolomite also occur as numerous veinlets. Quartz and chlorite veinlets are also present.

2.3.2 Pyrite

Pyrite (FeS_2) is the most abundant sulphide mineral present, accounting for approximately 35-40% of the total sulphides present. Pyrite grains range from euhedral to anhedral grains up to 1.5 mm across. The grains are often clustered to form massive sulphide blebs. Individual grains frequently contain inclusions of arsenopyrite, pyrrhotite, galena and chalcopyrite as well as a few mircograins of native gold (less than 0.05 mm). Many subhedral grains are highly fractured and may contain infillings of sulphosalts (jamesonite and boulangerite).

2.3.3 Arsenopyrite

Arsenopyrite (FeAsS) accounts for approximately 30% of the sulphide minerals present and is characteristically in the form of rhombohedral blades, 0.5 millimetres long. Inclusions of euhedral arsenopyrite occur in grains of pyrite, pyrrhotite and sphalerite.

2.3.4 Sphalerite

Sphalerite [(Zn,Fe)S] is locally abundant, constituting up to 10% of the metallic minerals. The grains are anhedral, individual grain boundaries are indistnct. Sphalerite grains host inclusions of euhedral pyrite, arsenopyrite, pyrrhotite, chalcopyrite and stannite. Sphalerite rich bands exist throughout the host rock.

2.3.5 Galena

Galena (PbS) forms large (2 mm) anhedral grains and may locally constitute up to 5% of the sulphide minerals. Galena may enclose inclusions of pyrrhotite, arsenopyrite and chalcopyrite. Grains of galena are present which are totally enclosed by pyrite.

2.3.6 Pyrrhotite

Pyrrhotite ($Fe_{1-x}S$) makes up a minor phase (less than 10%) of the sulphide mineralogy. Pyrrhotite laths up to 3 mm long are often clustered together. These large grains are usually extremely pitted and may be host to micro grains of native gold.

2.3.7 Chalcopyrite

Chalcopyrite ($CuFeS_2$) constitutes up to 4% of the total sulphides present. It occurs as auhedral grains up to 1 mm in diameter and occurs throughout as scattered blebs. Microprobe analysis identified the silver content in certain grains of chalcopyrite to be approximately 1000 ppm.

2.3.8 Rutile and Ilmenite

Rutile (TiO_2) and ilmenite ($TiFe_2O_5$) make-up approximately 3% of the metallic minerals. Skeletal ilmenite up to 1 mm long is common. In gangue, ilmenite pseudomorphs after feldspar grains are present. Rutile usually occurs with ilmenite as irregular masses of fine grained material (less than 0.01 mm).

2.3.9 Sulphosalts

X-ray diffraction analysis identified boulangerite

($Pb_5Sb_4S_{11}$) as the locally abundant sulphosalt. The microprobe, however, identified jamesonite ($Pb_4FeSb_6S_{14}$) as the major sulphosalt. Quantities of these two minerals appear to vary locally. Anhedral jamesonite is often observed sourrounding grains of euhedral pyrite.

2.3.10 Native Antimony and Allargentum

Native antimony and the alloy allargentum (3:1 ratio of Ag:Sb), identified by microscopy and microprobe work, appear to be closely related to the antimony-containing sulphosalts. Native antimony is found as small grains (less than 0.05 mm) with the sulphosalts.

2.3.11 Stannite and Cassiterite

Stannite (SnO_2) and cassiterite (Cu_2FeSnS_4) are usually found as small inclusions (less than 0.05 mm) in the sulphosalts and in sphalerite.

2.3.12 Native Gold

Native gold is widely dispersed. The gold occurs as micro grains (less than 0.01 mm) in contact with pyrite, pyrrhotite and arsenopyrite and sometimes as inclusions within these minerals. A minor number of large (0.15 mm) grains of gold have been noted. Native gold also occurs as free grains with the gangue.

2.4 Geochemistry

Drill core pulps from a section through the central part of the gabbro sill were systematically analyzed by Lacana (Hoy, 1988) in order to determine the variations in major and trace elements through the sill. The section included both mineralized and unmineralized gabbro.

The auriferous zone, comprising coarse grained, highly altered anorthositic gabbro, assayed 6.0 grams/tonne gold across 11 metres. The intrusive phases flanking the auriferous zones are comprised of moderately to strongly altered, fine to medium grained gabbro. Both upper and lower intrusive contacts are characterized by the development of cherty hornfels in the enclosing sediments.

The drill pulp analysis profile shows a strong, positively correlation of arsenic with the auriferous zone; arsenopyrite is the most abundant sulphide in the gold-bearing zones. Moderate, positive correlation of silica, iron and sodium are present whereas moderate, negative correlation of magnesium and calcium exist in the gold zone. Vanadium, cobalt, nickel and chromium are depleted in the gold zone.

The work completed has provided for a fairly detailed examination of the deposit along its strike, across its stratigraphic thickness and down its dip. The program of diamond drilling, surface trenching and sample analysis completed by Lacana to date indicates that within the area delineated as the "deposit" within the West Gabbro Zone, the mineralogy, including gold, varies very little.

Oxidation has affected the deposit to depths of several tens of metres, the result of permeability induced by the shearing the rocks have been subjected to. Oxidation affects the rocks along the shear and joint planes, the paths of least resistance to the movement of surface waters. Parallel to these planes, the rocks have developed a "rind" of oxidation products.

The intensity of the oxidation decreases with depth and is directly proportional to the degree of shearing the rock has been subject to.

3.0 Bulk Sampling

A bulk sample representing the Elmtree gold deposit was required in order to assess the various options available for processing Elmtree-type gold ores. Lacana had previously stripped and cleaned the unconsolidated overburden from the bedrock at several locations at the deposit site. A rock saw had been used to cut continuous channel samples for assay purposes across the geologic strike of the deposit at these exposures. One such exposure, trench #22, yielded an average of 6.86 grams/tonne of gold over its 10 metre length. This value compared favourably with the overall average gold content of the deposit as a whole, obtained from the sampling and assaying of extensive diamond drill core and several surface trenches. In addition, Mr. D. Hoy of Corona, who is intimately familiar with the geology of the deposit, indicated that not only is the gold content of the deposit homogeneous but the mineralogy of the deposit is also homogeneous. Based upon these factors, the material in trench #22 was believed to be representative of the deposit as a whole and therefore suitable for acquisition of the bulk sample. As a result, on September 27, 1988, the bulk sample was collected from trench #22.

3.1 Procedures

The showing is approximately 12 metres long and 5 metres wide with its long axis running perpendicular to the east-west strike of the deposit. An area within the showing, 3 metres long and 1 metre wide, with its long axis running parallel to that of the showing was chosen for the acquisition of the required bulk sample. Mr. David Doucette, owner of Green Thumb Landscaping, and aided by a labourer and a tailerable air compressor, drilled a 24 hole pattern (3 holes by 8 holes by 0.5 metres deep) within the 3 by 1 metre area to be sampled.

These holes were loaded with half a stick of 40% Forcite, tied-up with B-line primacord and shot using two E-B caps. A trench slightly greater than 3 by 1 metres and 1 metre deep was produced. The blasted rock was excavated using a tractor-mounted back-hoe subcontracted to Green Thumb. The waste rock piled to the west of the blasted trench. The sides of the trench were "scaled" of loose bedrock and the floor manually cleaned using a hand shovel. The naked air hose from the compressor was then utilized to clean the trench of all loose material.

A second area, approximately 2.5 metres long and 0.6 metres wide, on the floor of the blasted trench, was drilled on a 17 hole triangular pattern to a depth of 0.5 metres. The drilling rate for this second round of holes was appreciably slower than that for the first suggesting that the rock was more competent (less oxidized) at this depth. In addition, the rock dust produced from the drilling was distinctly grey in colour suggesting that the bulk of the rock being penetrated was fresh and unoxidized. This second round was blasted in a manner similar to the first and yielded a second, lower trench approximately 2.5 metres long by 0.6 metres wide by 0.75 metres deep. The blasted rock from this second, lower trench was taken for the bulk sample.

Four, clean 45-gallon, steel drums were lined with 6 mil construction plastic and filled with the broken rock from the lower (second) blast. As the rock was being loaded into the drums, grab samples were collected in a manner to represent the contents of each drum. Three samples representing each drum were collected. The plastic liners were then drawn up over the rock in each drum, tied and the lids secured.

The filled drums were trucked to Bathurst and off-loaded at the Armour Transport depot. On the return trip the loaded trucks were weighed on the scales at Elmtree Resources Ltd., the tare weights having been taken at the beginning of the day. It was determined that the four filled drums weighed 1,155 kilograms (includng the weight of the drums, approximately 15 kilograms each).

3.2 Sample Analyses

One set of the grab samples were delivered to Custom Lab in Bathurst and one set shipped to Chemlab in Saint John. Assays for gold were conducted by both labs; Custom analyzed two splits from their samples. ChemLab analyzed three splits. The results of the analyses were reported as follows (in grams per tonne):

	Custom Lab			ChemLab				overall
	split #1	split #2	average	split #1	split #2	split #3	average	average
drum #1	6.72	6.72	6.72	5.73	6.10	5.87	5.90	6.23
drum #2	4.53	4.25	4.39	5.65	6.58	6.87	6.37	5.58
drum #3	7.13	7.27	7.20	10.19	9.19	9.11	9.50	8.58
drum #4	5.38	5.35	5.37	4.98	5.02	5.00	5.00	5.15

The analytical results from Custom Lab were received within two days and were used, without ChemLab's results, to determine which drums were to be shipped to Lakefield Research. Since only half the rock collected was required by Lakefield, the two drums yielding neither the highest nor lowest gold values were shipped. Custom Lab's results indicated that the two drums to be shipped were #1 and #4 (6.72 and 5.37 grams/tonne, respectively). Subsequent receipt of ChemLab's analytical results, however, called this decision into question since they indicated that drums #1 and #4 contained 6.23 and 5.15 grams/tonne, respectively (overall average). Discussions with Robert Harris of NBDNRE indicated that they had need of only one drum of rock; as a result, drum #3 was shipped to NBDNRE in Fredericton (8.58

grams/tonne, overall average). Drum #2, originally thought to have the lowest gold content, was disposed of in Bathurst.

4.0 References

Hoy, D. 1988. The Elmtree deposit in The geologic setting of gold occurrences in Northern New Brunswick; New Brunswick Department of Natural Resources and New Brunswick Prospectors and Developers Association, Gold Symposium '88 Field Guide, p. 11-24.

Millar, R.E. 1988. Mineralogy of the Lacana Elmtree gold deposit, northern New Brunswick; unpublished B.Sc. Honours thesis, University of Waterloo, Waterloo, Ontario.

APPENDIX B

Preliminary Sizing Calculations
for
Oxidation Technologies

B.1.0 ACIDIC PRESSURE OXIDATION

Plant pre-design is based on the treatment of flotation concentrate and for the production of 75,000 troy oz/yr gold.

B.1.1 DESIGN CRITERIA

Conditions for pressure oxidation:

Temperature and pressure = 200°C, 100 psi*
Reaction time = 120 min*
Liquid solid ratio = 5/1 (15% w/w solids)*
Consumption of oxygen = 2.3 mt/mt sulfur in concentrate⁽¹⁾

Weight loss of solids in pressure oxidation = 6.8%*

Specific settling rate of solids for CCD = 1.23 m^2/tonne/day⁽¹⁾

Filtration rate = 0.31 m^2/tonne/day

Particle size requirements for oxidation: 97%, -400 mesh*

Conditions for cyanidation:

Concentration of sodium cyanide = 2 gpL, pH 10.5-11.0*
Residence time in CIP = 6 hrs*
Consumption of sodium cyanide = 7.48 kg/tonne*
Consumption of calcium oxide = 36.8 kg/tonne*
Recovery of gold = 96.1%

* bench test data (reagent consumption based on feed to oxidation).
(1) Sherritt Gordon approximation based on experience.

B.1.2 DRY SOLID THROUGHPUT CALCULATIONS

Production capacity of the plant = 75,000 troy oz/yr
Annual operating time = 8424 hr/yr (2 wk maintenance shut down)
Availability of flotation plant = 93.8%
Availability of oxidation plant = 95.8%
Availability of CIP plant = 97.9%
Recovery in flotation = 93%
Grade of ore = 5.71 g/tonne Au
Grade of concentrate = 24.4 g/tonne Au

$$\text{Annual tonnage of ore} = \frac{75{,}000\ (31.104)}{5.71\ (0.938)(0.958)(0.979)(0.93)(0.961)}$$

$$= 519{,}617.8 \text{ mt/yr}$$

For 351 days of operation: 1,480.4 mt/day (61.68 mt/hr)

$$\text{Feed to oxidation circuit} = \frac{519{,}617.8\ (5.71)(0.93)\ 24}{24.4\ (8424)}$$

$$= 322.2 \text{ mt/day}$$

$$\text{Feed to CIP circuit} = 322.2 - (322.2 \times 0.068)$$
$$= 300.3 \text{ mt/day}$$

B.1.3 EQUIPMENT SIZING CALCULATIONS

- based on lab testing and published papers
- acid pressure oxidation circuit as per drawing No. 4100-01-01

B.3.1.1 Regrind of Flotation Concentrate

Based on a work index of 13.3, power requirements are estimated as:

P = 80% passing 30 um for product
F = 80% passing 65 um for feed

$$W = \frac{10\ (13.3)}{\sqrt{30}} - \frac{10\ (13.3)}{\sqrt{65}}$$

$$= 7.79 \text{ kW·hr/st}$$

$$\begin{aligned} Hp &= 1.341\ (W)(t)(EF) \\ &= 1.341\ (7.79)(13.425)(0.9072)(1.2) \\ &= 153 \end{aligned}$$

Mill size selection based on Boliden-Allis Mills:

Overflow ball mill - 150 Hp
6′ diameter x 9′ length
(25 Hp/foot length)

B.1.3.2 Sizing of Pressure Autoclave Vessel

Size estimation based on L/d = 5.4 and 50% occupancy

Residence time = 2 hr (5 compartments, 6 agitators)

Solid throughput = 13.425 mt/hr

Reaction based on liquid solid ratio = 5/1 (15% solid)

$$\text{Volume of vessel} = \frac{13.425}{0.15} \times \frac{\text{m}^3}{1.132\ \text{mt}} \times 2\ \text{hr} \times 2$$

$$= 316.25\ \text{m}^3$$

$$\text{Vessel diameter} = \left[\frac{316.25\ (4)}{3.14\ (6.4)}\right]^{0.33}$$

$$= 3.97\ \text{m}$$

Vessel Length = 21.45 m

(Estimate based on each compartment operating at 15% solids, therefore, represents maximum volume)

$$\text{Agitation requirement} = 316.25\ \text{m}^3 \times 0.50 \times \frac{2.5\ \text{Hp}}{3.79\ \text{m}^3} \times 1.10\ \text{factor}$$

$$= 114.73\ \text{Hp total}$$

(20 Hp per agitator)

B.1.3.3 Oxygen Plant Sizing

Assumption: Consumption of oxygen is 2.3 mt O_2/mt sulfur
: 11-17% sulfur in concentrate, averaging 14%

Oxygen demand = 322.2 mt/day x 0.14 x 2.3
= 103.75 mt O_2/day
(78,435 m^3/day)

Sizing of oxygen plant basis:

- 100 mt O_2/day with an availability of 351 days per year
- will require a backup storage capacity in event of limited availability
- purity of oxygen - 95-98% O_2

B.1.3.4 Residence Washing - Counter Current Decantation (CCD)

Throughput tonnage of dry solids = 12.51 mt/hr

Specific settling rate = 1.23 m^2/mt/day (flocculant used)

Area = 12.51 mt/hr x 24 hr/day x 1.23
= 369.36 m^2

Thickener diameter = 21.69 m (71.2 ft)

For heavy duty thickener:

$$Hp = \frac{150{,}000 \times 2\ RPM}{5250 \times 50\%\ EFF}$$

$$= 1.14$$

B.2.0 BIO-OXIDATION

Plant pre-design is based on the treatment of flotation concentrate and for the production of 75,000 troy oz/yr gold.

B.2.1 DESIGN CRITERIA

Conditions for bio-leach:

Temperature = 35°C*
Uptake of oxygen = 68 mg O_2/L/hr (0.27 m^3/m^3/hr)*
Heat input by oxidation = 920 kcal/kg solid$^{(1)}$
Reaction rate = based on batch tests, see section B.2.3*
Specific settling rate of solids = 1.23 m^2/tonne/day
Weight loss of solids from oxidation = 20%*

Conditions for cyanidation:

Consumption of sodium cyanide = 7.53 kg/t*
Consumption of calcium oxide = 17.80 kg/t*
Recovery of gold = 90.4% at 69.6% extraction of iron*
Concentration of sodium cyanide = 2 gpL, pH 11.0*
Residence time in CIP = 24 hrs*

* bench test data (reagent consumption based on oxidation feed)
(1) pilot data of Giant Bay Biotech Inc.

B.2.2 DRY SOLID THROUGHPUT CALCULATIONS

NOTE: The basis for throughput calculations are similar to pressure leach (as outlined in Section B.1.2), except for:

Availability of the oxidation plant = 97.9%
Recovery of gold in oxidation and CIP circuit = 90.4%

$$\text{Annual tonnage of ore} = \frac{75{,}000\ (31.104)}{5.71\ (0.938)(0.979)(0.979)(0.93)(0.904)}$$

$$= 540{,}532.5\ mt/yr$$

For 351 days of operation: 1,540.0 mt/day (64.6 mt/hr)

$$\text{Feed to oxidation circuit} = \frac{540{,}532.5\ (5.71)(0.93)\ 24}{24.4\ (8424)}$$

$$= 335.2 \text{ mt/day}$$

$$\text{Feed to CIP circuit} = 335.2 - (335.2 \times 0.20)$$
$$= 268.2 \text{ mt/day}$$

B.2.3 EQUIPMENT SIZING CALCULATIONS

- based on lab testing and published papers
- regrind of concentrate, similar to size outlined in Section B.1.3.1
- biological oxidation as per drawing No. 4100-02-02

B.2.3.1 Sizing of the Bio-Leach Circuit

Based on retention time - iron extraction relationships of test program.

NOTE: Regeneration of bio-mass and the maximum extraction of iron occurs between 6-15 days of batch leach.

From the data (of Table No. 10 of the Lakefield Report), Time vs. Iron Extraction, the extraction rate is 4.763% Fe/day based on a correlation coefficient of 0.970. The maximum extraction rate (omitting data at time 9,11,14 days) is 4.981% Fe/day based on a correlation coefficient of 0.985.

Assuming an extraction of 69.6% Fe for a gold recovery of 90.4%, the residence time is expressed by:

$$t = \frac{69.6\%\ \text{Fe}}{4.981\%\ \text{Fe/day}}$$

$$= 13.87 \text{ days } (333 \text{ hrs})$$

In order to optimize on throughput, a gold recovery of 84.0% requires an extraction of 67.1% Fe and the residence time requirements would be 323 hr. The residence time or leach volume will decrease by 3%. However, the throughput for design production capacity would increase to 360.7 mt/day or by 8%. This would mean a net increase in capacity of 5%.

The sizing of the leach circuit will be based on 333 hrs of residence time.

Slurry density at 15% solids = 1.117 mt/m^3 (based on a solid sp.gr. of 3.32)

$$\text{Slurry flowrate} = \frac{335.2 \text{ mt/day solid}}{24\ (0.15)(1.117 \text{ mt/m}^3)}$$

$= 83.358\ m^3$/hr slurry (13.967 mt/hr dry solids)

Total volume of leach circuit = 83.358 (333)
= 27,758 m^3

Selection of tank size to be approximately 12.195 m (40') diameter 12.195 height.

Tank volume = 1,423.7 m^3
Total number of tanks = 20

Agitation requirements are based on mechanical agitation and air rate.

Assumption:

For bio-leach systems, data has not been published on effects of mixer power and gas rates. D/T ratios (impellor to tank diameter) from 0.35-0.45 are necessary to provide blending and heat transfer, and to prevent physical shear to the organisms. These ratios are not most effective for gas-liquid mass transfer. For high D/T ratios, power requirements are reduced and torque increases. The ratio of mixer power/gas rate is unknown. It is assumed that power requirements are as follows (these estimations will have to be confirmed by a pilot plant program).

Superficial gas velocity = 0.13 m/sec

Gas mass transfer coefficient = apprx. 0.2 kg mole /(hr)(m^3)(bar)

Power level = 0.12 kW/m^3 (0.6 Hp/1000 gal)

Power requirements per tank = 1,423.7 m^3 (0.12 kW/m^3)
= 170.8 kW (233 Hp)

B.2.3.2 Compressor Sizing for Bio-Leach Demand

Discharge pressure = 207 kPa Absolute (30 psia)

Utilization factor of 50% (2.0)

Demand = 0.270 m^3 air/m^3/hr x 1427.7 m^3 x 2.0 x 2.0 tanks
= 15,419 m^3/hr (9,066 cfm @ STP)

From Power Handbook 2nd Ed. pg. 70:

For an adiabatic compression factor of 1.2

A single stage compressor 1000 cfm = 48 Hp

For 15,419 m^3/hr, Hp = 435

For a compression ratio of 2, the efficiency = 68%

Total power requirements = 640 Hp (468 kW)

B.2.3.3 Heat Balance in Bio-Leach Circuit

NOTE: Heat load is dependent on the sulfide content.

For concentrates containing 11% S, 6.44% As, and 23.9% Fe:

Throughput = 13.967 mt/hr solids

Heat input from reaction at steady state = 920 Kcal/kg solid

Reaction maintained at 35°C

	SUMMER	WINTER
Outside Temperature	25°C	-25°C
Slurry Temperature to Bio-Leach	25°C	25°C
Net Heat Input by Oxidation (Kcal/hr)	-12,849,333	-12,849,333
Heat Loss Estimate from Action	10%	35%
Heat Loss (Kcal/hr)	+1,284,933	+4,497,267
Net Heat Input (Kcal/hr)	-11,564,400	-8,352,066
	13.450 mmW	

For cooling water entering cooling tower at 31°C and exit at 26°C, the net cooling water flowrate is based on a 95% efficiency of heat exchange:

	SUMMER	WINTER
Cooling water flow (m^3/hr)	2,434 (0.676 m^3/s)	1,758

Estimated heat transfer area based on an overall heat transfer coefficient of 6,000 W/m^2 °C and 5°C change in temperature.

Area = 448 m^2

B.2.3.4 Counter Current Decantation Circuit

Similar to CCD for pressure leach Section B.1.3.4

Feed to CIP = 268.2 mt/day (11.175 mt/hr)

Thickener area = 329.89 m^2

Diameter = 20.5 m (67.3 ft)

B.3.0 ROASTING

- Equipment sizing is based on standard industrial practice.

B.3.1 DESIGN CRITERIA

CONDITIONS FOR ROASTING	STAGE ONE	STAGE TWO
Reactor Fluidization (m/sec)	0.10	0.085
Operating Temperature °C	550*	650*
Free Board Temperature °C	530	580 lwr 440 upr
Bed Diameter to Depth Ratio	2.582	1.830
Residence Time	3 hr	1.5 hr
Bed Cooling Water (L/mt feed)	72	72

Weight loss of solids in roasting = 16%* (10% from 1° stage)

Calcine specific settling rate of solids for CCD = 1.23 m^3/tonne/day

Particle size requirements for oxidation = 97%, -400 mesh*

Consumption of sodium cyanide = 2.42 kg/t*
Consumption of lime = 11.6 kg/t*

* bench scale data (reagent consumption based on feed to oxidation)

B.3.2 DRY SOLID THROUGHPUT CALCULATIONS

NOTE: The basis for throughput calculations are similar to pressure leach (as outlined in Section B.1.2), except for:

Availability of the roaster plant = 91.7%
Recovery of gold in oxidation and CIP circuit = 79.2%

$$\text{Annual tonnage of ore} = \frac{75{,}000\ (31.104)}{5.71\ (0.938)(0.917)(0.979)(0.93)(0.792)}$$

$$= 658{,}685.95\ \text{mt/yr}$$

For 351 days of operation: 1,876.6 mt/day (78.2 mt/hr)

$$\text{Feed to oxidation circuit} = \frac{658{,}685.95\ (5.71)(0.93)\ 24}{24.4\ (8424)}$$

$$= 408.4\ \text{mt/day}$$

Feed to CIP circuit = 408.4 - (408.4 x 0.16)
= 343.1 mt/day

B.3.3 EQUIPMENT SIZING CALCULATIONS

- Based on standard operating practice by Dorr-Oliver and lab test results.
- Fluidized bed roasting as per drawing No. 4100-03-03.

B.3.3.1 Roaster Sizing

For a feed rate of 17.02 mt/hr (3.32 mt/m^3 bulk density):

Volume of bed with 69.2 void = 49.93 m^3 for primary
= 22.47 m^3 for secondary

For a bed diameter to depth ratio of:

2.582 for primary - bed depth = 2.177 m
- bed diameter = 5.466 m (17.93 ft)

1.830 for secondary - bed depth = 1.987 m
- bed diameter = 3.635 m (11.9 ft)

Volume of fluidization air:

For primary (based on 0.10 m/s) = 8,444 m^3/hr
(4,965 cfm)

For secondary (based on 0.085 m/s) = 3,176 m^3/hr
(1,868 cfm)

Fan sizing for fluidization air:
(Power requirements 8.8 kW/1000 m^3/hr)

For primary = 43.69 kW·hr at 100% efficiency
= 67.21 kW·hr at 65% efficiency

For secondary = 43.00 kW·hr at 65% efficiency
(60 Hp)

Ratio of air for primary to secondary = 2.66:1

B.3.3.2 Flue Gas Volume and Sulfuric Acid Production Rate

Temperature of gas before scrubbing = 102°C (cooled to sublime arsenic trioxide)

Free board gas volume = 11,620 N m^3/hr

NOTE: Various assumptions have been made to simplify calculations. The density of the off-gas is about 5.0×10^{-4} g/cm^3 at 550°C and has a molecular weight of approximately 26 g/g·mole. The gas composition is as follows (percentage by mass):

63% N_2
19% H_2O
15% SO_2
3% O_2

Volumetric flow of gas at 550°C = 11,620 x 925°K/375°K
= 28,663 m^3/hr (28,663 x 10^6 cm^3/hr)

Mass flow of gas at 550°C = 551,205 g·mole/hr

Heat capacity calculations based on gas composition and heat capacity of components.

	Temperature	Cp N_2	Cp H_2O	Cp SO_2	Cp O_2	Cp (gas)
			(cal/mole°C)			
Cooling Air	298°K	6.96	8.02	9.54	7.02	7.42
Scrubber Gas	375°K	7.04	8.19	10.32	7.25	7.59
Freeboard Gas	925°K	7.60	9.55	11.17	8.04	8.38

Enthalpy for reference of 273°K (20°C)

h_{298} = 7.42 cal/g·mole°C (25-20°C)
= 37.1 cal/g·mole

h_{375} = 7.59 cal/g·mole°C (102-20°C)
= 622 cal/g·mole

h_{823} = 8.38 cal/g·mole°C (550-20°C)
= 4,441 cal/g·mole

Amount of heat given off by the cooling of freeboard gas:

$Q = m(H_2 - H_1)$
= 551,205 g·mole/hr (4,441-622 cal/g·mole)
= 2.105×10^9 cal/hr

Amount of cooling air required to cool freeboard gas:

$M = Q/(H_2 - H_1)$
= $2.105 \cdot 10^9$ cal/hr / (622-37.1 cal/g·mole)
= 3,598,756 g·mole/hr
= 93.567×10^6 g/hr

At a gas density of $1.55 \cdot 10^{-3}$ g/cm^3, gas volumetric flow is:

= 93.567×10^6 g/hr / 1.55×10^{-3} g/cm^3 $\times 10^{-6}$ m^3/cm^3
= 60,366 N m^3/hr

Ratio of cooling air to fluidization air = 5.2 (volumetric basis)

Ratio of cooling air to fluidization air at Campbell Red Lake Mine = 4.1

NOTE: Calculations are based on no heat loss from reactors and that the temperature of the before cooling is 550°C. At the Campbell Red Lake Mines gold roaster in Balmeston, Ontario, the gas temperature from the roaster cyclones is 400°C.

Total gas flow to scrubber is approximately:

= 60,366 + 8,444 + 3,176 N m^3/hr
= 72,008 N m^3/hr (42,340 cfm)

Estimated concentration of SO_2 = 2.8% (wt/wt)

Tonnage of SO_2 = $72,008 \times 10^6$ cm^3/hr $\times 1.55 \times 10^{-3}$ g/cm^3
$\times 0.028 \times$ mt/10^6 g

= 3.13 mt/hr

Check on mass balance of sulfur

Assumption: Removal of sulfur from concentrate is 100% efficient

Amount of sulfur in feed = 408.4 mt/day / 24 hr/day x 0.11 S/solids
= 1.87 mt/hr sulfur
= 3.74 mt/hr SO_2

NOTE: This is based on minimum sulfur loading using low grade concentrate as feedstock. Maximum sulfur loading may be up to 5.79 mt/hr SO_2 for 17% S in concentrate. Maximum concentration of flue gas would be 5.2% SO_2.

B.3.3.3 Calcine Regrind

Power requirements similar to regrind of flotation of concentrate (see B.1.3.1)

W = 7.79 kW·hr/st

Hp = 1.341 (7.79)(14.3 mt/hr)(0.9072) 1.2
= 162.6

Mill sizing is similar to milling required for bio and pressure leach.

B.3.3.4 Counter Current Wash

Feed to CIP = 14.3 mt/hr

As per calculations in section B.1.3.4

Area = 14.3 (24)(1.23)
= 422.14 m^2

Thickener diameter = 23.2 m (76 ft)

APPENDIX C

Preliminary Sizing Calculations for Primary Upgrading

C.1.0 DESIGN CRITERIA

Crushing Specifications: To treat run of mine ore (by primary, secondary and tertiary crushing) to reduce particle size to 80%, -12.7 mm.

Grinding Specifications: To treat crushed ore (by rod mill and ball mill grinding) to reduce particle size to 85%, -74 microns, work index of 13.3 kW•hr/st (14.7 kW•hr/mt).

Flotation Specifications: Consists of a bulk flotation to produce a single concentrate by rougher, scavenger and cleaner circuits.

Rougher/Scavenger: residence time 0.38 hr
Cleaner: residence time 0.10 hr

Reagent Consumption:	200 g/t	AX-350
	10 g/t	M2030
	30 g/t	MIBC
	250 g/t	$CuSO_4$
	45 g/t	R-2D8

Estimated grade/recovery from rougher = 16 g/t at 86%

Additional grade/recovery from scavenger and cleaner circuits

Overall grade/recovery = 24.4 g/t at 93%

C.2.0 EQUIPMENT SIZE ESTIMATIONS

	ACIDIC PRESSURE LEACH	BIO-LEACH	ROASTING
Throughput of ore (mt/day)	1,480.4	1,540.0	1,876.6
Estimated Power for Crushing (kW)			
Primary	35.5	36.9	49.3
Secondary	63.6	66.2	80.6
Tertiary	72.5	75.4	91.9
Total kW•hr/mt	2.8	2.8	2.8
Estimated Power for Regrinding (kW)			
Rod Mill	255.8	266.1	324.3
Ball Mill	795.5	829.7	1,010.9
Rod Mill Size (dia x length m)	2.7 x 4.1	2.7 x 4.1	2.7 x 4.1
Ball Mill Size (dia x length m)	3.7 x 5.5	3.7 x 5.5	3.9 x 5.9
Total kW•hr/mt	16.9	16.9	16.9
Rougher/Scavenger Cell Operation			
Throughput (mt/hr)	61.68	64.17	78.19
% Solids in Feed*	34	34	34
Volumetric Feed Rate (m^3/hr)	144.4	150.3	183.1
Total Cell Volume (m^3)	54.9	57.1	69.6
No. of Cells (3 m^3)	18	19	23
Cleaner Cell Operation			
Throughput (mt/hr)	11.89	12.39	15.10
% Solids in Feed**	8	8	8
Volumetric Feed Rate (m^3/hr)	139.7	145.6	177.4
Total Cell Volume (m^3)	13.9	14.6	17.7
No. of Cells (1 m^3)	14	15	18

* solid sp.gr. 2.5
** solid sp.gr. 4.0

APPENDIX D

Preliminary Sizing Calculations
for
Cyanide Leaching and Carbon in Pulp Circuit

D.1.0 DESIGN CRITERIA

Plant design is based on the treatment of refractory gold flotation concentrate requiring oxidation prior to total cyanidation.

Design Criteria:

Average circuit feed rat = 11.8 mt/hr(1)
Cyanide leach time = 36 hrs
CIP stages = 6
CIP residence time = 6 hrs
Carbon loading = 5,000 g/mt (eluted value of 50 g/mt carbon)
Carbon concentration in solution = 20 kg/m^3

(1) circuit feed rate for each technology is as follows (for 75,000 oz/yr):

pressure leach = 12.51 mt/hr
bio-leach = 11.17 mt/hr
roasting = 14.29 mt/hr

- Average feed rate = 11.8 mt/hr for bio-leach and pressure oxidation having a sizing error of ± 6%.
- CIP for roasting is considered 21% larger than average sizing.

C.2.0 EQUIPMENT SIZING

- The cyanide leach and carbon in pulp gold recovery circuit as per drawing 5100-01/02 - 04/05
- The reactivation and striping of carbon and the electrowinning-refining to produce a gold dore.

C.2.1 CYANIDE LEACH CIRCUIT

Slurry density at 55% solids: 1.75 mt/m^3

Slurry flowrate = 11.8 mt/hr + m^3/1.75 mt
= 12.26 m^3/hr

Total tank volume = 12.26 m^3/hr x 36 hr
= 441 m^3

$$\text{Size of each tank} = \frac{\pi d^2}{4} (1.15\ d) = 0.9032\ d^3 = \frac{441\ m^3}{6\ \text{tanks}}$$

d = 4.33

Select 4.33 m diameter x 4.98 m slurry height

Tank size with freeboard = 4.33 m diameter x 5.24 m

Horsepower requirements on agitators = $\frac{73.56\ m^3}{tank} \times \frac{0.625\ Hp}{m^3}$

= 46 Hp

D.2.2 REPULP AND ALKALINE CONTROL TANKS

Residence time per tank = 0.5 hr

Total tank volume = 12.26 m^3 x 0.5 hr
= 6.13 m^3

Size of each tank = 0.9032 d^3 = 6.13 m^3
d = 1.89 m

Select 1.89 m diameter x 2.17 m slurry height

Tank size with freeboard = 1.89 m diameter x 2.38 m

Horsepower requirements on agitators = $\frac{6.13\ m^3}{tank} \times \frac{0.625\ Hp}{m^3}$

= 3.8 Hp

D.2.3 CIP TRASH SCREENS

Screens are required to prevent entry of wood chips/plastic into CIP circuit

CIP screen of 20 mesh in circuit

Trash screen size selected at 28 mesh

Standard high speed wire cloth screen 4′ x 8′

D.2.4 CIP HOLDING TANK

The holding tank is to provide a surge capacity to allow continuous operation in event of down-time in the cyanidation circuit or CIP circuit.

Residence time in tank = 12 hr

Total volume of tank = 12.26 m^3/hr x 12 hr
= 147.12 m^3

Size of tank = 0.9032 d^3 = 147.12 m^3
d = 5.45 m

Select 5.45 m diameter x 6.27 slurry height

Size of tank with freeboard = 5.45 m diameter x 6.58 m

Horsepower requirements on agitators = 147.12 m^3 x $\frac{0.635\ Hp}{m^3}$

= 94 Hp

D.2.5 CIP SLURRY FEED PUMPS

Design CIP to operate at 45% solids to allow for dilution by carbon washing

Maximum slurry density in feed pump is 43% solid, 1.503 mt/m^3

Volumetric flowrate = $\frac{10.03\ mt/hr}{0.43}$ x $\frac{m^3}{1.503\ mt}$

= 15.52 m^3/hr (68.3 US gpm)

Assume head to be about 60 H-ft based on slurry density

Pump size = 1.5" x 2.5", 10 Hp (9" runner)

D.2.6 CARBON IN PULP TANKS

Design slurry density is 43% solids, total residence time of 7 hrs.

Design flow at 15.52 m^3/hr at 1 hour/stage requires 15.52 m^3 tank

Size of tank = 0.9032 d^3 = 15.52 m^3
d = 2.58 m

Select 2.58 m diameter x 2.97 slurry height

Size of tank with freeboard = 2.58 m diameter x 3.12 m

Horsepower requirements on agitators = $15.52\ m^3 \times \frac{0.635\ Hp}{m^3}$

$= 10\ Hp$

D.2.7 CIP INTERSTAGE SCREENS

Use air blown peripheral screens ... lap screens

Total screen area per tank

5 panels x 1.05 m long x 0.4 m = 2.1 m^2

Total screen requirement at 20 mesh (7 tanks) = 14.7 m^2

D.2.8 CARBON ADVANCE IN CIP TANKS

Design bases for carbon loading is 5,000 g Au/mt carbon.

NOTE: The rate of loading for carbon may be approximately 1,000 g of Au per hour and will load to approximately 7,000 g Au/mt carbon based on a 7 hour residence time in CIP. At 7,000 g Au/mt, the transfer rate will be less, therefore, design at 5,000 g Au/mt may be considered conservative for design of carbon regeneration.

Required carbon movement = $\frac{10.03\ mt/hr \times 24\ hr/day \times 30.56\ g/mt}{5{,}000\ g\ Au/mt\ carbon}$

= 1.4713 mt carbon/day
= 61.303 kg/hr (1,471.27 kg/day)

Carbon concentration in solution = 20 kg carbon/m^3 liquid

Volume of liquid per tank = $(1 - 0.43) \times 15.52\ m^3 \times \frac{1.503\ mt}{m^3} \times \frac{m^3}{1.09\ mt\ liquid}$

= 12.198 m^3 liquid/tank

Carbon inventory in CIP = $12.198\ m^3 \times \frac{20\ kg\ carbon}{m^3} \times 7$ tanks

= 1,707.76 kg

Carbon concentration in slurry = $\frac{1{,}707.7\ kg}{7(15.52\ m^3)} = 15.719\ \frac{kg\ carbon}{m^3\ slurry}$

Slurry transfer rate set at 22.73 m^3/hr (100 US gpm) to avoid flooding of loaded carbon screens.

$$\text{Required slurry movement} = 1{,}471.27\ \frac{kg}{day} \times \frac{m^3\ slurry}{15.719\ kg\ carbon}$$

$$= 93.60\ m^3\ slurry/day$$

$$\text{For slurry transfer, the total time required} = \frac{93.60\ m^3\ slurry/day}{22.73\ m^3/hr}$$

$$= 4.12\ hr/day$$

For two transfers per day, total time per transfer = 2 hr

Transfer pump size with a vertical spindel = 2½" x 2½"

D.2.9 CIP TAILING THICKENER AND PROCESS WATER TANK

Minimum slurry density is 40% solids, 1.452 mt/m^3

$$\text{Flowrate} = \frac{10.03\ mt/hr}{0.40} \times \frac{m^3}{1.42\ mt} = 17.274\ m^3/hr$$

Specific settling rate = 1.23 $m^2/mtpd$

$$\text{Area} = 1.23\ m^2 \times 10.03\ mt/hr \times 24\ hr/day$$
$$= 296.09\ m^2$$

Diameter of thickener = 19.42 m (63.70 ft)

Overflow of water = 15.045 m^3/hr

Overflow rate = 1.21 $m^3/day/m^2$ (29.86 US gal/day/ft^2)

Water rise velocity = 0.0508 m/hr

For heavy duty thickener, torque on rake is approximately 150,000 ft·lb

$$Hp = \frac{150{,}000 \times 2\ RPM}{5250 \times 50\%\ efficiency}$$

$$= 1.14$$

Thickener underflow at 70% solids, 2.195 mt/m^3

$$\text{Volumetric flowrate} = \frac{10.03\ mt/hr}{0.70} \times \frac{1\ m^3}{2.195\ mt}$$

$$= 6.528\ m^3/hr\ \ (28.72\ US\ gpm)$$

Pump size = 3" ODS

Process water tank, holding 2 hr supply of water

Tank size = 15.045 m^3/hr x 2 hr = 30.1 m^3

Tank diameter = 3.2 m

Height = 3.69 m

Process water pump flowrate = 15.045 m^3/hr (66.2 US gpm)

Pump size for 60' head = 2" x 2"

D.2.10 CARBON STRIPPING VESSELS - PRESSURE STRIP

Conditions for pressure strip = 100 psig at 150°C for 6-8 hrs

Carbon movement = 1471.3 kg/day

Bulk density of dry carbon = 0.47 t/m^3

Design basis = 3 batch/day, 16 hr/batch, 2 batch reactors

Design tonnage = 490.43 kg/batch

Design batch volume = 1 m^3 (bed volume)

Based on a reactor design of 60° cone and height of freeboard above cone to diameter is 1.21:1. Reactor design pressure is 1.5 operating pressure.

Based on a volume requirement for washing, having a strip solution flow of 12 bed volumes for current industrial operating practice, using 0.1% NaOH and 0.1% NaCN solution, and 2 bed volumes for wash:

Total volume of strip solution = 36 m^3/day
Total volume of wash solution = 6 m^3/day

Total solution to electrowin = 42 m^3/day (1.75 m^2/hr)

Storage for electrowin solution to allow for continuous electrowinning design for 0.8 daily capacity:

Volume of tank = 33 m^3

Diameter approximately = 3.2 m

D.2.11 ELECTROWIN CELLS/INDUCTION FURNACE

Based on a continuous operation, flow to cell = 1.75 m^3/hr

For a 1 hour residence time, cell size = 1.75 m^3

Amount of steel wool = 75,000 oz/yr gold x 0.454 $\frac{\text{kg steel wool}}{\text{100 oz gold}}$ x 1 yr/351 days

= 0.97 kg/day

Size of induction furnace = 150 kW for 1,200°C max operation

D.2.12 CARBON REGENERATION KILN

Operation at 600-800°C for 12-20 min

Size based on carbon movement rate = 1.5 mt/day

Length of kiln = 1.5 mt/day x 0.747 m/mt/day
= 1.12 m

Diameter = 0.12 m

Gas heating rate = 1.5 mt/day x 0.634 mm BTU/hr/mt/day
= 1 mm BTU/hr

APPENDIX E

Preliminary Sizing Calculations for Tailings Treatment

E.1.0 DESIGN CRITERIA

Operating conditions for neutralization of oxidation liquors:

Flotation tailing may be added to 1st stage as a pH modifier

First stage pH adjustment to 5 = 33.5 kg $CaCO_3$ equivalent/tonne slurry*

Second stage pH adjustment to 11 = 11.8 kg CaO equivalent/tonne slurry*

Residence time for each stage = 40 min

Solid precipitation = 0.2 kg solid/ kg solution*

Operating conditions for cyanidation solid treatment:

Total cyanide concentration on a monthly average = 100 mg/L

Sulfur dioxide = 5 kg/kg CN_T

Copper sulfate ($CuSO_4 \cdot 5\,H_2O$) = 0.16 kg/kg CN_T

Lime = 7 kg/kg CN_T

Solids in pulp = 40% solids

Residence time = 1 hr

Air requirements = 18 m^3/m^3 slurry

Specific settling rate of tailing = 0.5 m^2/mt/day

* based on neutralization data from lab program.

pH 5 ... 24.8 kg/t $Ca(OH)_2$
pH from 5 ... 15.6 kg/t $Ca(OH)_2$
(no tailing addition)

E.2.0 PRELIMINARY EQUIPMENT SIZING

- Based on commercial installation operating data.
- Tailings treatment circuit as per drawing No. 6100-01-06.

E.2.1 NEUTRALIZATION OF OXIDATION LIQUORS (BIO-LEACH AND PRESSURE OXIDATION)

Volumetric flow of oxidation solution based on 15% solids

Solution volume = (13.417 mt/hr ÷ 0.15) - 13.417 mt/hr ÷ 1.06 mt/m^3
= 71.73 m^3/hr (76.03 mt/hr)

Total residence time of circuit = 1.3 hr

Individual tank volume = 71.73 m^3/hr x 1.3 hr ÷ 4
= 23.3 m^3

Diameter of each tank = 3.09 m

Power requirements for agitation = 0.5 kW/m^3 x 23.3 m^3
= 11.65 kW (15 Hp)

E.2.2 CYANIDE REMOVAL CIRCUIT

Total tonnage of solids = 12.5 mt/hr

At 45% solids, volumetric throughput = 12.5 mt/hr ÷ 0.40 ÷ 1.393 mt/m^3
= 22.43 m^3/hr

Volume of each tank = 22.43 (1) (0.33)
= 7.5 m^3

Diameter = 2.1 m

Power requirements for agitation = 0.5 kW/m^3 x 7.5 m^3
= 3.75 kW (5 Hp)

E.2.3 TAILING THICKENER

Total solids to tails = Flotation Tails + CN Residue + Neut PPt.
= 48.25 mt/hr + 12.5 mt/hr + 0.2 (76 mt/hr)
= 75.95 mt/hr

Thickener area = 75.95 mt/hr x 0.5 m^2/mt/day x 24
= 911 m^2

Diameter = 34 m

APPENDIX F

Capital Cost Estimates

F.1.0 BASIS FOR COSTING

Costs based on 3rd quarter 1988 pricing:

Marshall & Swift Equipment Cost Index (M&S):	856.5
Mining and Milling = 873.9	
Chemical Engineering Plant Cost Index (CE):	348.3
Equipment:	380.7
Construction Labour:	268.6
Buildings:	324.4
Engineering and Supervision:	342.8

Canadian currency relative to American: $US = 1.20 $CDN

Relative costing for crushing, grinding, flotation, carbon in pulp, counter current decantation circuits.

Budget costing done on pressure leach, oxygen, roaster, and bio-leach circuits.

Installation costs are based on equipment supplier recommendations or factored (based on industrial practice).

Electrical, instrumentation and piping costs are factored based on the installation of comminution, beneficiation and hydrometallurgical operations.

Building, civil works, yard improvement, and land is based on actual size requirements.

Engineering, supervision and construction are based on a factor of total direct costs.

Level of confidence for preliminary budget costing based on lab results is -5%/+30%.

F.2.0 EQUIPMENT COSTS

F.2.1 CRUSHING AND ORE STORAGE

Based on USBM IC 9143, capital cost includes the acquisition and installation of equipment. Includes primary, secondary and tertiary crusher, screens, and the attendant materials handling equipment such as feeders, belt conveyors etc. Costs have been adjusted based on M&S 1984 index of 780.4. The crushing operation is based on 3 shifts per day with coarse ore storage of 4,500 mt.

	PRESS. OXIDATION P.OX.	BIO-OXIDATION B.OX.	ROASTING FBR
Throughput of Ore (mt/day)	1,480.39	1,539.98	1,876.59
Purchase Equipment Cost	1,297,600.00	1,318,000.00	1,427,000.00
Construction Supply & Labour Cost	493,500.00	501,000.00	543,000.00
Transportation Cost	36,500.00	37,000.00	40,000.00
Total	1,827,600.00	1,856,000.00	2,010,000.00

F.2.2 PRIMARY AND SECONDARY GRINDING

Based on USBM IC 9143, capital cost includes rod mill and ball mill with the major associated equipment such as feeders, conveyors, pumps and classifiers. The ore hardness factor is 1.019 based on a work index of 14.7 kW·hr/mt. The size factor is 1.658 based on feed and product size requirements.

	P.OX.	B.OX.	FBR
Purchase Equipment	2,477,000	2,539,000	2,977,000
Construction Supply & Labour Cost	1,029,000	1,067,000	1,251,000
Transportation Cost	71,000	74,000	87,000
Total	3,547,000	3,680,000	4,315,000

F.2.3 FLOTATION FOR MEDIUM TO LARGE SIZE PLANT

Based on USBM IC 9143, the equipment purchase includes conditions, pumps, flotation cells (for self aerating type) for rougher scavenger and cleaner cells.

	P.OX.	B.OX.	FBR
Purchase Equipment	455,600	460,000	483,000
Construction Supply & Labour	213,900	216,000	227,000
Transportation	20,800	21,000	22,000
Total	690,300	697,000	732,000

F.2.4 CONCENTRATE THICKENING

Based USBM IC 9143, the equipment purchase includes earth work construction of tank, walk-way pumps drive mechanism and rake. Setting area adjustment factor for 0.50 m^2/mt/day is 0.649.

	P.OX.	B.OX.	FBR
Flotation Concentrate (mt/day)	322.2	335.2	408.4
MSRL Factor	2.67	2.69	2.78
Purchase Cost	213,000.00	218,000.00	254,000.00
Installation Cost	244,000.00	251,000.00	293,000.00
Transportation Cost	4,000.00	4,000.00	5,000.00
Total	461,000.00	473,000.00	552,000.00

MSRL: mild steel rubber lined

F.2.5 COUNTER CURRENT DECANTATION (CCD), OXIDATION SOLID WASHING

Based on USBM IC 9143, the CCD includes thickener mechanism and pumps for 2-stage circuit. Adjustment factor for 1.23 m^2/mt day, settling rate is 1.60. Thickeners tanks of mild steel.

	P.OX.	B.OX.	FBR
Oxidation Product Tonnage (mt/day)	299.6	268.1	343.1
Purchase Cost	375,000.00	349,000.00	408,000.00
Installation Cost & Transp.	440,000.00	410,000.00	478,000.00
Total	815,000.00	759,000.00	886,000.00

F.2.6 CCD SOLID FILTRATION, DRUM FILTER

Based on USBM IC 9143, filtration includes drum filter, wet-type vacuum pumps, filtrate pumps, slurry pumps, air blowers, and filtrate receivers. The filtration rate factor is 1.

	P.OX.	B.OX.	FBR
Purchase Cost	206,000	191,000	224,000
Installation Cost & Transp.	101,000	94,000	111,000
Total	307,000	285,000	335,000

F.2.7 CARBON IN PULP (CIP) CIRCUIT

Based on USBM IC 9143, costing for CIP includes pumps, carbon inventory, thickening, leaching, screens, carbon adsorption, counter current transfer, zadra-type stripping, acid treatment, reactivation of carbon by heating, electrowinning, acid treatment, scavenger recovery from bleed streams, bullion refining and casting.

	P.OX.	B.OX.	FBR
Purchase Cost	1,330,000	1,241,000	1,445,000
Installation Cost	2,469,000	2,306,000	2,685,000
Total	3,799,000	3,547,000	4,130,000

F.2.8 OXIDATION SOLUTION NEUTRALIZATION

(For pressure oxidation and bio-oxidation solution treatment only)

Based on Moncton Foundry and Machinery and Prochem agitators.

Four neutralization tanks C.S. 7122 HAR Placite Lined, 3 m d x 3 m h

Purchase Cost:	42,000

Four agitators, rubber lined, 12 kW

Purchase Cost:	60,000
Total Purchase Cost:	102,000
Installation:	52,000
Total	154,000

F.2.9 CYANIDE REMOVAL CIRCUIT

(For all three technologies)

Costs based on Moncton Foundry and Machinery (MFM) and Prochem agitators.

Three reaction tanks C.S. 7122 HAR Placite Lined, 2 m d x 2 m h

Purchase Cost:	16,000

Three agitators, 3.75 kW

Purchase Cost:	11,000

Air Blower (100 cfm)

Purchase Cost:	4,000

Total Purchase Cost:	31,000
Installation:	15,000
Total	46,000

F.2.10 TAILING THICKENER

Based on USBM IC 9143, cost of thickener, drive mechanism, rake, and pumps.

For 1822 mt/day of solids and a settling area adjustment factor of 0.65.

Purchase equipment cost:	235,000
Installation:	276,000
Total	551,000

E.2.11 LIME AND LIMESTONE

Based on USBM IC 9143, lime slaking includes dry storage, ball mill slaking, cyclone classification, and slurry storage.

	P.OX.	B.OX.	FBR
Lime Demand(1) (mt/day)	33.5	27.1	4.0
Limestone Demand(2) (mt/day)	68.4	68.4	-
(Net lime demand for acid-gas scrubbing [using 85% efficiency] would be 92.4 mt/day CaO, having an installed cost of $907,000.)			
Lime:			
Purchase Equipment Cost	286,000	258,000	104,000
Installation Cost	274,000	248,000	100,000
Total	560,000	506,000	204,000
Limestone:			
Purchase Equipment Cost	401,000	401,000	-
Installation Cost	385,000	385,000	-
Total	786,000	786,000	-
Subtotal	1,346,000	1,292,000	204,000

(1)(2) See operating cost estimates of Appendix G for calculation of reagent consumption.

Lime contains 98% CaO
Limestone contains 50% CaO equivalent

F.2.12 PRESSURE LEACH OXIDATION EQUIPMENT

Cost is based on discussions with Sherritt Gordon External Technology. Cost includes autoclave (with titanium compounds and lead brick lining), flash tanks, regrind circuit, and slurry storage.

Total Capital Installed Cost:	16,000,000
Purchase Cost:	12,000,000
Installation Cost:	4,000,000

F.2.13 BIO-LEACH OXIDATION CIRCUIT

The total installation cost is based on budget costing of compounds of major capital expense.

Tanks fabrication done by MFM (Moncton, New Brunswick) using mild carbon steel, coated with placite (7122 HAR) for protection from abrasion and insulated. A complexity factor of 1.2 has been used to adjust costing for manholes, walk-way and flanged connections for heat exchange hookups.

20 tanks (12.2 m dia x 12.2 m H)

Purchase Cost:	4,823,000
Installation and Transportation:	2,411,000
Total:	7,234,000

Agitators for tanks based on discussion with Prochem, maximum horsepower requirements for 250 Hp motor, drive and agitator.

20 agitators

Purchase cost:	2,500,000
Installation and Transportation:	500,000
Total:	3,000,000

Cooling tower based on Chem. Eng. Magazine/Nov. 1988. Cost covers the erected cost for modular type cooling towers, fans, pumps, and motors.

For 2,434 m^3/hr water throughput

Installation and Purchase Cost:	78,000
Transportation:	3,000
Total:	81,000

Total Installed Cost for Bio-Leach Circuit: 10,315,000

F.2.14 FLUIDIZED BED ROASTER

Based on discussions with Dorr Oliver Canada Ltd., budget pricing for a gold roasting system using a two-stage reactor configuration with burner, blowers, cyclones, refractory, precipitator, ducting, controls, feeder, quench tank, and baghouse. A budget price for 10.25 mt/hr is approximately 3.8 mm Canadian dollars. A power factor of 0.70 is assumed for purchase cost.

For 17 mt/day:

Purchase Cost :	5,500,000
Installation (50%):	2,750,000
Transportation (3%):	165,000
Engineering (15%):	825,000
Total:	9,240,000

F.2.15 SULFURIC ACID PLANT

Based on cost projections for conventional acid plants (Bechtel Canada Inc., Aug., 1988), the capital cost based on battery limits for a double contact plant sized for the production of 580 mt/day sulfuric acid (from a gas volume of 67,000 N m^3/hr at 10.5% SO_2), having a direct cost of 21.3 mm Canadian dollars. For the production of 140 mt H_2SO_4 per day, a power factor of 0.70 is assumed for total direct cost. Cost includes storage and loadout facility, having a scrubbing efficiency of 99%.

Total Direct Cost:	7,870,000
Purchase Cost:	5,250,000
Installation Cost:	2,620,000

NOTE: Equivalent to direct capital investments of $56,000/tonne/day.

F.2.16 REGRIND MILLS CIRCUIT

Based on budget costing from Sala Machine Works and a power factor of 0.70 based on ball mill horsepower.

	B.OX.	FBR
Purchase Cost:	178,000	186,000
Installation Cost:	89,000	93,000
Total	267,000	279,000

F.3.0 INSTRUMENTATION, PIPING, ELECTRICAL, AND PLANT SERVICES

The following factors have been used to cost materials and labour based on the purchase cost of equipment. The assumptions have been based on assumed industrial requirement for each technology.

	P.OX.	B.OX.	FBR
Instrumentation	15%	10%	15%
Piping	20%	25%	10%
Electrical	20%	22%	15%
Services/Utilities	10%	10%	10%

F.4.0 BUILDING/YARD AND LAND

The net requirement for each technology has been based on a pilot plant and unit costs for civil works etc.

	P.OX.	B.OX.	FBR
Process Building Area (m^2)	3,296	2,656	2,656/640
Height (m)	13.72	13.72	13.72*
Unit Cost ($/$m^2$)	915	915	915/1,022
Utility Building Area (m^2)	105	105	105
Height (m)	4.6	4.6	4.6
Unit Cost ($/$m^2$)	750	750	750
Total Cost (Building)	3,095,000	2,509,000	3,163,000
Site Area/Fence (m^2/m)	15,000/500	19,000/1,000	15,000/500
Unit Cost ($/$m^2$)/($/m)	75/50	75/50	75/50
Preparation Cost (yard)	1,150,000	1,475,000	1,150,000
Total Land Requirements			
Tailing Area* (km^2)	10	10	2
Unit Cost ($/$km^2$)	12,000	12,000	12,000
Land Purchase Cost	120,000	120,000	24,000

* area housing roaster and acid plant has a total height of 18.3 m (areas includes 512 m^2)

** based on tailing area requirements for 20 yrs (80% less land is required for roaster option).

APPENDIX G

Operating and Maintenance Cost Estimates

F.1.0 BASIS FOR COSTING

Costs based on 3rd quarter 1988 pricing:

- Wages for employees are based on standard industrial practice for metallurgical and chemical processing within New Brunswick. Overtime, burden, and fringe benefits are factored.

- Reagent costs are based on published prices in the Chemical Marketing Reporter (copyright by Schnell Publishing Company Inc.). If reagents can be purchased locally, pricing will be based on local supply budget pricing.

- Operating and maintenance costs have been factored based on similar size operations. For each oxidation technology, operating and maintenance costs are based on published literature and standard industrial practices.

- Costs for refining of dore could not be obtained without prior knowledge of dore composition, frequency of shipment and tonnages. It is assumed that at 1989 metal credit would be 99%.

- Value of feedstock (ore) to the custom mill has been based on a mining capital cost of 30 million Canadian dollars and an operating cost of 15 Canadian dollars per tonne of ore.

- Plant operation is based on three 12 hour shifts per day, 351 days per year with a two week shut down for maintenance overhauls.

G.2.0 OPERATING AND MAINTENANCE LABOUR COSTS

NOTE: Labour costs do not include administration, and are based on standard industrial wage rates for New Brunswick.

ASSUMPTION: Staffing requirements for each technology would be similar. Staffing for oxygen and sulfuric acid plant would be included in operating cost for oxygen and acid.

	TITLE OCCUPATION	ANNUAL SALARY/WAGE $CDN X 1000	NUMBER REQUIRED	TOTAL DOLLARS PER YEAR $CDN X 1000
Supervision Staff	Plant Superintendent	70	1	70
	Assistant Superintendent	60	1	60
	General Mill Foreman	50	2	100
	Engineers	50	4	200
	Chemist/Assayers	40	2	80
	Lab Technicans	25	5	125
	Production Clerk	20	1	20
	Security	30	4	120
Operating Staff	Shift Foreman	45	4	180
	Control Room	40	8	320
	Crushing	35	8	280
	Grinding/Flotation	35	8	280
	Oxidation/Thickeners	35	8	280
	Leach Circuit	35	4	140
	Carbon Treatment	35	8	280
	Tailing Treatment/Reagents	30	4	120
	Shift Assistants/Labourers	25	8	200
	Sampler	25	4	100
Maintenance Staff	(day shift only)			
	Elect/Mech/Inst Foreman	50	1	50
	Electricians	35	4	140
	Millwrights	35	2	70
	Mechanics	35	4	140
	Machinist, Welder	35	2	70
	Instrumentation Technicians	35	3	105
	Greasers/Labourers	25	2	50
	Total Staff		98	
	Total Annual Salary	(33.24)		3,580
	Overtime (10%)			358
	Burden/Fringe Benefits (32%)			1,146
	Net Total Annual Salary			5,084

G.3.0 REAGENT CONSUMPTION AND ANNUAL COSTS

NOTE: Cost for reagents used in flotation have been included in operation cost calculations for flotation. The reagent costs have been based on consumptions identified in the bench test program.

G.3.1 LIME

Cost for lime from a New Brunswick source is about $94/mt, FOB Sussex, N.B.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Consumption (mt/day)			
CIP	11.05	4.77	3.98
Neutralization	21.50	21.50	0.00
CN Removal	0.32	0.32	0.32
Acid Plant Effluent*	-	-	0.20
Total	32.82	26.59	4.50
Annual Purchase Cost	1,083,000.00	877,000.00	142,000.00
Shipping (350 km)	472,000.00	382,000.00	62,000.00
Total	1,554,000.00	1,259,000.00	204,000.00

* effluent based on 0.07 kg/hr H_2SO_4/mt/day acid produced (production of acid is 140.3 mt/day)

G.3.2 LIMESTONE

Cost for lime from a New Brunswick source is about $17/mt, FOB Bathurst, N.B.

ASSUMPTION: CAO equivalent of 50%

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Consumption (mt/day)			
Neutralization	68.00	68.00	0.00
Annual Purchase Cost	406,000.00	406,000.00	0.00
Shipping (100 km)	279,000.00	279,000.00	0.00
Total	685,000.00	685,000.00	0.00

G.3.3 SODIUM CYANIDE

Cost for sodium cyanide is based on $CDN 1.98/kg, FOB Montreal, Quebec.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Consumption (mt/day)	2,410.00	2,524.00	988.00
Annual Purchase Cost	1,675,000.00	1,754,000.00	690,000.00
Shipping (400 km)	40,000.00	41,000.00	16,000.00
Total	2,075,000.00	1,795,000.00	706,000.00

G.3.4 OXYGEN FOR PRESSURE LEACH OXIDATION

Cost based on over-the-fence purchase from a Linde oxygen plant to be owned and operated by Linde and budget costing for terms of 15 years is as follows:

For 110,000 scfm oxygen at 200 psi, 95% O_2 minimum.

Purchase Cost:	196,700/month
Annual Cost:	2,360,400 ($CDN 67.25/mt O_2)

G.3.5 FLOCCULANT

Price based on costs of 3,130/mt FOB Montreal, Quebec.

Normal addition rates are based on 3 mt/L of polymer to thickener feed slurries. Operating costs are similar for each technology.

Flocculant Consumption (kg/day)	
Flotation Concentrate Thickener:	18.6
CCD Circuit (2 Thickeners):	8.0
CIP Thickener:	2.6
Tailing Thickener:	4.6

Purchase Cost:	37,100
Transportation (400 km):	500
Total:	37,600

G.3.6 INCO'S SULFUR DIOXIDE - AIR CYANIDE REMOVAL PROCESS

Cost based on Inco's cost for a 1,000 mt/day operation and a power factor of 1.

Reagent consumption based on cost per mt solids.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Sulfuric Dioxide ($0.63/mt)*	67,000	59,000	76,000
Copper Sulfate ($0.08/mt)	9,000	8,000	10,000
Lime ($0.35/mt)	37,000	33,000	42,000
Royalty ($0.82/mt)	87,000	77,000	99,000
Total Cost	200,000	177,000	227,000

* cost of liquid SO_2 based on $250/mt at site
costs are calculated based on feed tonnage to CIP

G.3.7 FLOTATION REAGENTS

Cost based on USBM IC 9143, for operating costs. Reagent usage is based on the flotation of copper ore and includes pH adjustment, frother, and collectors. A power factor of one is used for reagents.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Total	570,000	592,000	722,000

G.3.8 NUTRIENTS FOR BIO-LEACH

Studies done by P.M. Mineral Leaching Technologies Inc. in Burnaby, B.C., Canada, the total cost of nutrient salts based on continuous piloting is as follows:

CONSUMPTION	TOTAL COST
$4.19/mt	493,000

The composition of the nutrients is not reported, however, is assumed to be similar to the 9 K mixture report by Lakefield.

G.4.0 VALUE OF ORE AT MINE SITE

The value of the ore has been based on projected mining costs for underground mining in the Bathurst area. The cost would depend on depth and nature of the orebody. Based on average costs for 1,000-1,500 mt/day operations, the mining operation and maintenance is based on $CDN20/mt and a net capital worth of $CDN30 million. Assuming a 15% ROI before taxes for 15 years the net value of the run of mine is $CDN29.87/mt.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Annual Ore Cost	15,521,000	16,146,000	19,676,000

G.5.0 ACID PLANT

For the production of sulfuric acid from roaster off-gas, the total operating costs including labour, power supplies and maintenance have been based on published literature. It has been estimated that the operating and maintenance cost is about $CDN40/mt of sulfuric acid (98%). Net production of sulfuric acid based on 3.74 mt SO_2/hr is about 140,3 mt/day (49,000 mt/yr).

Total O&M Cost: 1,970,000/yr

Revenue from acid based on $50/mt FOB Bathurst is approximately $2,450,000/yr.

G.6.0 UTILITY COSTS

G.6.1 ELECTRICITY

For crushing, grinding, flotation, and CIP, the consumption of electricity has been factored based on USBM IC 9143. The consumption of electricity for oxidation and tailing treatment has been based on estimate power requirements for equipment.

Power rates are based on $CDN 0.0425/kW·hr.

POWER REQUIREMENTS kW	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Ore Tonnage (mt/day)	1480.4	1539.4	1,876.59
(mt/hr)	61.68	64.14	78.19
Crushing	159	165	194
Grinding	947	983	1,194
Flotation	243	253	309
Oxidation	193	2,922*	298
CIP	763	741	800
Tailing Treatment	82	82	82
Utilities (air, etc.)	150	150	150
Total	2,542	5,295	3,027
Total Annual Cost	910,000	1,896,000	1,084,000

* equivalent to $889/mt concentrate

G.7.0 MAINTENANCE/OPERATING SUPPLIES AND INSURANCE

ASSUMPTION: Maintenance cost will be based on 3% of direct cost and operating supplies will be 20% of labour cost.

Insurance cost is based on $0.35/100 direct cost.

	PRESS. OXIDATION	BIO-OXIDATION	F.B. ROASTING
Maintenance Cost	1,301,000	1,064,000	1,246,000
Operating Supplies	1,017,000	1,017,000	1,017,000
Insurance	152,000	124,000	145,000
Total	2,470,000	2,205,000	2,408,000

G.8.0 REFINING CHARGE

Based on a credit of 98% from refinery.

For 75,000 oz/yr valued at $CDN479/oz.

Refining Charge = $718,500

APPENDIX H

Financial Results

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                    ***          OUTPUT LISTING            ***
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                         Economic Analysis - Constant CDN Dollars ('000, 1989)
                              Custom Mill - Biological Oxidation Technology
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
ANNUAL EXPENSES											
Concentrate		22887	22887	22887	22887	22887	22887	22887	22887	22887	22887
Labour		3633	3633	3633	3633	3633	3633	3633	3633	3633	3633
Reagents		4431	4431	4431	4431	4431	4431	4431	4431	4431	4431
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		1227	1227	1227	1227	1227	1227	1227	1227	1227	1227
Supplies/Insurance		1585	1585	1585	1585	1585	1585	1585	1585	1585	1585
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		34482	34482	34482	34482	34482	34482	34482	34482	34482	34482
CAPITAL OUTLAYS											
Purchase of Equipment		10687	0	0	0	0	0	0	0	0	0
Installation cost		6789	0	0	0	0	0	0	0	0	0
Instrumentation		1069	0	0	0	0	0	0	0	0	0
Piping		2672	0	0	0	0	0	0	0	0	0
Electrical		2351	0	0	0	0	0	0	0	0	0
Plant Services		1069	0	0	0	0	0	0	0	0	0
Building		885	0	0	0	0	0	0	0	0	0
Land		72	0	0	0	0	0	0	0	0	0
Engineering		5119	0	0	0	0	0	0	0	0	0
Construction indirects		768	0	0	0	0	0	0	0	0	0
Commissioning/Spares		512	0	0	0	0	0	0	0	0	0
Contingency		3839	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		3583	0	0	0	0	0	0	0	0	0
SUBTOTAL		39415	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		73897	34482	34482	34482	34482	34482	34482	34482	34482	34482
CASH FLOW		-37897	1518	1518	1518	1518	1518	1518	1518	1518	1518
ACCUMULATED CASH FLOW		-37897	-36379	-34861	-33343	-31825	-30307	-28789	-27271	-25753	-24235

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000
ANNUAL EXPENSES						
Concentrate		22887	22887	22887	22887	22887
Labour		3633	3633	3633	3633	3633
Reagents		4431	4431	4431	4431	4431
Acid Plant		0	0	0	0	0
Electricity		1227	1227	1227	1227	1227
Supplies/Insurance		1585	1585	1585	1585	1585
Refining Charge		719	719	719	719	719
SUBTOTAL		34482	34482	34482	34482	34482
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0	0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		34482	34482	34482	34482	34482
CASH FLOW		1518	1518	1518	1518	1518
ACCUMULATED CASH FLOW		-22717	-21199	-19681	-18163	-16645

ECONOMIC RATE OF RETURN	-6.88
PAYBACK PERIOD	> 16.00
TERMINAL WORTH	36

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
TOTAL		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
ANNUAL EXPENSES											
Concentrate		343305	299963	264645	235621	211573	191488	174586	160256	148022	137507
Labour		54495	47615	42009	37402	33584	30396	27713	25439	23497	21827
Reagents		66465	58074	51236	45617	40961	37073	33800	31026	28658	26622
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		18405	16081	14188	12632	11343	10266	9360	8592	7936	7372
Supplies/Insurance		23775	20773	18328	16318	14652	13261	12091	11098	10251	9523
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		517230	451930	398720	354992	318760	288500	263035	241445	223013	207170
CAPITAL OUTLAYS											
Purchase of Equipment		10687	10687	10687	10687	10687	10687	10687	10687	10687	10687
Installation cost		6789	6789	6789	6789	6789	6789	6789	6789	6789	6789
Instrumentation		1069	1069	1069	1069	1069	1069	1069	1069	1069	1069
Piping		2672	2672	2672	2672	2672	2672	2672	2672	2672	2672
Electrical		2351	2351	2351	2351	2351	2351	2351	2351	2351	2351
Plant Services		1069	1069	1069	1069	1069	1069	1069	1069	1069	1069
Building		885	885	885	885	885	885	885	885	885	885
Land		72	72	72	72	72	72	72	72	72	72
Engineering		5119	5119	5119	5119	5119	5119	5119	5119	5119	5119
Construction indirects		768	768	768	768	768	768	768	768	768	768
Commissioning/Spares		512	512	512	512	512	512	512	512	512	512
Contingency		3839	3839	3839	3839	3839	3839	3839	3839	3839	3839
WORKING CAPITAL		3583	3583	3583	3583	3583	3583	3583	3583	3583	3583
TERMINAL WORTH		-36	-27	-21	-16	-12	-9	-7	-6	-5	-4
SUBTOTAL		39379	39388	39394	39399	39403	39406	39408	39409	39410	39411
TOTAL ANNUAL COSTS		556609	491318	438114	394391	358163	327906	302442	280855	262424	246581
NET PRESENT WORTH:		-16609	-19492	-21841	-23771	-25370	-26705	-27829	-28780	-29593	-30291
BENEFIT COST RATIO		0.97	0.96	0.95	0.94	0.93	0.92	0.91	0.90	0.89	0.88

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

SENSITIVITY ANALYSIS

OPERATING COSTS

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Concentrate		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	42.76	18.56	-INF	-INF
TERMINAL WORTH	$	36	36	36	36
PAYBACK PERIOD	YEARS	0.32	2.89	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.35	1.13	0.85	0.76
AT 2 %		1.33	1.11	0.84	0.75
AT 4 %		1.31	1.10	0.84	0.75
AT 6 %		1.29	1.09	0.83	0.74
AT 8 %		1.27	1.07	0.82	0.73
AT 10 %		1.25	1.06	0.81	0.73
AT 12 %		1.23	1.04	0.80	0.72
AT 14 %		1.21	1.03	0.80	0.71
AT 16 %		1.19	1.02	0.79	0.71
AT 18 %		1.17	1.00	0.78	0.70

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

SENSITIVITY ANALYSIS		REVENUES			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Gold Metal		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-INF	-INF	19.54	45.21
TERMINAL WORTH	$	36	36	36	36
PAYBACK PERIOD	YEARS	16.00	16.00	2.70	0.20
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.68	0.82	1.12	1.26
AT 2 %		0.67	0.82	1.10	1.25
AT 4 %		0.67	0.81	1.09	1.24
AT 6 %		0.66	0.80	1.08	1.22
AT 8 %		0.65	0.79	1.07	1.21
AT 10 %		0.64	0.78	1.06	1.19
AT 12 %		0.64	0.77	1.04	1.18
AT 14 %		0.63	0.76	1.03	1.17
AT 16 %		0.62	0.75	1.02	1.15
AT 18 %		0.61	0.75	1.01	1.14

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Biological Oxidation Technology

SENSITIVITY ANALYSIS — CAPITAL COST

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-3.08	-5.19	-8.27	-9.45
TERMINAL WORTH	$	25	31	41	47
PAYBACK PERIOD	YEARS	16.00	16.00	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.99	0.98	0.96	0.95
AT 2 %		0.98	0.97	0.95	0.94
AT 4 %		0.97	0.96	0.94	0.93
AT 6 %		0.97	0.95	0.93	0.91
AT 8 %		0.96	0.94	0.92	0.90
AT 10 %		0.95	0.93	0.90	0.89
AT 12 %		0.94	0.92	0.89	0.88
AT 14 %		0.93	0.92	0.88	0.86
AT 16 %		0.93	0.91	0.87	0.85
AT 18 %		0.92	0.90	0.86	0.84

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                             ***          OUTPUT LISTING            ***
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                                  Integrated Plant - Biological Oxidation Technology
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Biological Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
ANNUAL EXPENSES											
Ore Purchase		16146	16146	16146	16146	16146	16146	16146	16146	16146	16146
Labour		5084	5084	5084	5084	5084	5084	5084	5084	5084	5084
Reagents		5001	5001	5001	5001	5001	5001	5001	5001	5001	5001
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		1896	1896	1896	1896	1896	1896	1896	1896	1896	1896
Supplies/Insurance		2205	2205	2205	2205	2205	2205	2205	2205	2205	2205
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		31051	31051	31051	31051	31051	31051	31051	31051	31051	31051
CAPITAL OUTLAYS											
Purchase of Equipment		14922	0	0	0	0	0	0	0	0	0
Installation cost		8960	0	0	0	0	0	0	0	0	0
Instrumentation		1492	0	0	0	0	0	0	0	0	0
Piping		3730	0	0	0	0	0	0	0	0	0
Electrical		3283	0	0	0	0	0	0	0	0	0
Plant Services		1492	0	0	0	0	0	0	0	0	0
Building		1475	0	0	0	0	0	0	0	0	0
Land		120	0	0	0	0	0	0	0	0	0
Engineering		7095	0	0	0	0	0	0	0	0	0
Construction indirects		1064	0	0	0	0	0	0	0	0	0
Commissioning/Spares		709	0	0	0	0	0	0	0	0	0
Contingency		3547	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		4789	0	0	0	0	0	0	0	0	0
SUBTOTAL		52678	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		83729	31051	31051	31051	31051	31051	31051	31051	31051	31051
CASH FLOW		-47729	4949	4949	4949	4949	4949	4949	4949	4949	4949
ACCUMULATED CASH FLOW		-47729	-42780	-37831	-32882	-27933	-22984	-18035	-13086	-8137	-3188

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Biological Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000
ANNUAL EXPENSES						
Ore Purchase		16146	16146	16146	16146	16146
Labour		5084	5084	5084	5084	5084
Reagents		5001	5001	5001	5001	5001
Acid Plant		0	0	0	0	0
Electricity		1896	1896	1896	1896	1896
Supplies/Insurance		2205	2205	2205	2205	2205
Refining Charge		719	719	719	719	719
SUBTOTAL		31051	31051	31051	31051	31051
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0	0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		31051	31051	31051	31051	31051
CASH FLOW		4949	4949	4949	4949	4949
ACCUMULATED CASH FLOW		1761	6710	11659	16608	21557

ECONOMIC RATE OF RETURN	5.41
PAYBACK PERIOD	7.64
TERMINAL WORTH	60

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Biological Oxidation Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
TOTAL		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
ANNUAL EXPENSES											
Ore Purchase		242190	211614	186698	166223	149257	135089	123164	113055	104425	97006
Labour		76260	66632	58787	52340	46998	42536	38782	35599	32881	30545
Reagents		75015	65544	57827	51485	46230	41842	38148	35017	32344	30046
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		28440	24849	21924	19519	17527	15863	14463	13276	12262	11391
Supplies/Insurance		33075	28899	25497	22700	20384	18449	16820	15440	14261	13248
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		465765	406962	359047	319670	287043	259794	236862	217421	200823	186556
CAPITAL OUTLAYS											
Purchase of Equipment		14922	14922	14922	14922	14922	14922	14922	14922	14922	14922
Installation cost		8960	8960	8960	8960	8960	8960	8960	8960	8960	8960
Instrumentation		1492	1492	1492	1492	1492	1492	1492	1492	1492	1492
Piping		3730	3730	3730	3730	3730	3730	3730	3730	3730	3730
Electrical		3283	3283	3283	3283	3283	3283	3283	3283	3283	3283
Plant Services		1492	1492	1492	1492	1492	1492	1492	1492	1492	1492
Building		1475	1475	1475	1475	1475	1475	1475	1475	1475	1475
Land		120	120	120	120	120	120	120	120	120	120
Engineering		7095	7095	7095	7095	7095	7095	7095	7095	7095	7095
Construction indirects		1064	1064	1064	1064	1064	1064	1064	1064	1064	1064
Commissioning/Spares		709	709	709	709	709	709	709	709	709	709
Contingency		3547	3547	3547	3547	3547	3547	3547	3547	3547	3547
WORKING CAPITAL		4789	4789	4789	4789	4789	4789	4789	4789	4789	4789
TERMINAL WORTH		-60	-45	-35	-27	-20	-16	-12	-10	-8	-6
SUBTOTAL		52518	52633	52643	52651	52658	52662	52666	52668	52670	52672
TOTAL ANNUAL COSTS		518383	459595	411690	372321	339700	312456	289528	270090	253494	239228
NET PRESENT WORTH:		21617	12230	4583	-1702	-6909	-11255	-14914	-18015	-20663	-22938
BENEFIT COST RATIO		1.04	1.03	1.01	1.00	0.98	0.96	0.95	0.93	0.92	0.90

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Biological Oxidation Technology

SENSITIVITY ANALYSIS		OPERATING COSTS			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Ore Purchase		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	36.66	20.70	-23.29	-INF
TERMINAL WORTH	$	50	50	50	50
PAYBACK PERIOD	YEARS	0.69	2.48	16.00	16.00

DISCOUNT RATES

BENEFIT COST RATIO

	1	2	3	4
AT 0 %	1.43	1.20	0.92	0.82
AT 2 %	1.40	1.18	0.91	0.81
AT 4 %	1.37	1.16	0.89	0.80
AT 6 %	1.34	1.14	0.88	0.79
AT 8 %	1.31	1.12	0.87	0.78
AT 10 %	1.28	1.10	0.86	0.77
AT 12 %	1.26	1.08	0.84	0.76
AT 14 %	1.23	1.06	0.83	0.75
AT 16 %	1.20	1.04	0.82	0.74
AT 18 %	1.18	1.02	0.81	0.73

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                                       Refractory Gold Processing Options
                              Economic Analysis - Constant CDN Dollars ('000, 1989)
                               Integrated Plant - Biological Oxidation Technology
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SENSITIVITY ANALYSIS                  REVENUES
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ANALYSIS No.                             1        2        3        4
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DESCRIPTION                  L/F      % CHANGE % CHANGE % CHANGE % CHANGE
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Gold Metal                             -30.00   -15.00    15.00    30.00


ECONOMIC RATE OF RETURN  %            -INF     -INF      23.12    42.34
TERMINAL WORTH           $               60       60       60       60
PAYBACK PERIOD           YEARS        16.00    16.00     2.09     0.34
DISCOUNT RATES
--------------
BENEFIT COST RATIO

    AT  0  %                           0.73     0.89     1.20     1.35
    AT  2  %                           0.72     0.87     1.18     1.33
    AT  4  %                           0.71     0.86     1.16     1.31
    AT  6  %                           0.70     0.85     1.14     1.29
    AT  8  %                           0.69     0.83     1.13     1.27
    AT  10  %                          0.67     0.82     1.11     1.25
    AT  12  %                          0.66     0.81     1.09     1.23
    AT  14  %                          0.65     0.79     1.07     1.21
    AT  16  %                          0.64     0.78     1.06     1.19
    AT  18  %                          0.63     0.77     1.04     1.13
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Biological Oxidation Technology

SENSITIVITY ANALYSIS — CAPITAL COST

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	11.67	8.11	3.27	1.51
TERMINAL WORTH	$	42	51	69	78
PAYBACK PERIOD	YEARS	4.74	6.19	9.10	10.55
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.07	1.06	1.03	1.01
AT 2 %		1.06	1.04	1.01	1.00
AT 4 %		1.05	1.03	0.99	0.98
AT 6 %		1.04	1.02	0.98	0.96
AT 8 %		1.02	1.00	0.96	0.94
AT 10 %		1.01	0.99	0.94	0.92
AT 12 %		1.00	0.97	0.93	0.90
AT 14 %		0.99	0.96	0.91	0.89
AT 16 %		0.97	0.95	0.89	0.87
AT 18 %		0.96	0.93	0.88	0.85

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DESCRIPTION :
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                                          Refractory Gold Processing Options
                                 Economic Analysis - Constant CDN Dollars ('000, 1989)
                                  Custom Mill - Acidic Pressure Oxidation Technology
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
ANNUAL EXPENSES											
Concentrate		22102	22102	22102	22102	22102	22102	22102	22102	22102	22102
Labour		3633	3633	3633	3633	3633	3633	3633	3633	3633	3633
Reagents		6874	6874	6874	6874	6874	6874	6874	6874	6874	6874
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		427	427	427	427	427	427	427	427	427	427
Supplies/Insurance		1848	1848	1848	1848	1848	1848	1848	1848	1848	1848
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		35603	35603	35603	35603	35603	35603	35603	35603	35603	35603
CAPITAL OUTLAYS											
Purchase of Equipment		14966	0	0	0	0	0	0	0	0	0
Installation cost		8012	0	0	0	0	0	0	0	0	0
Instrumentation		2245	0	0	0	0	0	0	0	0	0
Piping		2993	0	0	0	0	0	0	0	0	0
Electrical		2993	0	0	0	0	0	0	0	0	0
Plant Services		1497	0	0	0	0	0	0	0	0	0
Building		690	0	0	0	0	0	0	0	0	0
Land		72	0	0	0	0	0	0	0	0	0
Engineering		6694	0	0	0	0	0	0	0	0	0
Construction indirects		1004	0	0	0	0	0	0	0	0	0
Commissioning/Spares		669	0	0	0	0	0	0	0	0	0
Contingency		4686	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		5020	0	0	0	0	0	0	0	0	0
SUBTOTAL		51541	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		87144	35603	35603	35603	35603	35603	35603	35603	35603	35603
CASH FLOW		-51144	397	397	397	397	397	397	397	397	397
ACCUMULATED CASH FLOW		-51144	-50747	-50350	-49953	-49556	-49159	-48762	-48365	-47968	-47571

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000
ANNUAL EXPENSES						
Concentrate		22102	22102	22102	22102	22102
Labour		3633	3633	3633	3633	3633
Reagents		6874	6874	6874	6874	6874
Acid Plant		0	0	0	0	0
Electricity		427	427	427	427	427
Supplies/Insurance		1848	1848	1848	1848	1848
Refining Charge		719	719	719	719	719
SUBTOTAL		35603	35603	35603	35603	35603
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0	0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		35603	35603	35603	35603	35603
CASH FLOW		397	397	397	397	397
ACCUMULATED CASH FLOW		-47174	-46777	-46380	-45983	-45586

ECONOMIC RATE OF RETURN	-21.11
PAYBACK PERIOD	> 16.00
TERMINAL WORTH	36

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
TOTAL		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
ANNUAL EXPENSES											
Concentrate		331530	289674	255568	227540	204316	184921	168598	154760	142945	132790
Labour		54495	47615	42009	37402	33584	30396	27713	25439	23497	21827
Reagents		103110	90092	79485	70768	63545	57513	52436	48132	44458	41299
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		6405	5596	4937	4396	3947	3573	3257	2990	2762	2565
Supplies/Insurance		27720	24220	21369	19025	17083	15462	14097	12940	11952	11103
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		534045	466622	411682	366532	329123	297879	271586	249295	230264	213905
CAPITAL OUTLAYS											
Purchase of Equipment		14966	14966	14966	14966	14966	14966	14966	14966	14966	14966
Installation cost		8012	8012	8012	8012	8012	8012	8012	8012	8012	8012
Instrumentation		2245	2245	2245	2245	2245	2245	2245	2245	2245	2245
Piping		2993	2993	2993	2993	2993	2993	2993	2993	2993	2993
Electrical		2993	2993	2993	2993	2993	2993	2993	2993	2993	2993
Plant Services		1497	1497	1497	1497	1497	1497	1497	1497	1497	1497
Building		690	690	690	690	690	690	690	690	690	690
Land		72	72	72	72	72	72	72	72	72	72
Engineering		6694	6694	6694	6694	6694	6694	6694	6694	6694	6694
Construction indirects		1004	1004	1004	1004	1004	1004	1004	1004	1004	1004
Commissioning/Spares		669	669	669	669	669	669	669	669	669	669
Contingency		4686	4686	4686	4686	4686	4686	4686	4686	4686	4686
WORKING CAPITAL		5020	5020	5020	5020	5020	5020	5020	5020	5020	5020
TERMINAL WORTH		-36	-27	-21	-16	-12	-9	-7	-6	-5	-4
SUBTOTAL		51505	51514	51520	51525	51529	51532	51534	51535	51536	51537
TOTAL ANNUAL COSTS		585550	518136	463202	418058	380651	349411	323119	300830	281800	265442
NET PRESENT WORTH:		-45550	-46311	-46930	-47438	-47859	-48210	-48505	-48755	-48969	-49152
BENEFIT COST RATIO		0.92	0.91	0.90	0.89	0.87	0.86	0.85	0.84	0.83	0.81

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS — OPERATING COSTS

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Concentrate		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	26.34	8.56	-INF	-INF
TERMINAL WORTH	$	36	36	36	36
PAYBACK PERIOD	YEARS	1.65	5.98	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.27	1.07	0.81	0.72
AT 2 %		1.25	1.05	0.80	0.72
AT 4 %		1.23	1.04	0.79	0.71
AT 6 %		1.20	1.02	0.78	0.70
AT 8 %		1.18	1.00	0.77	0.69
AT 10 %		1.16	0.99	0.76	0.69
AT 12 %		1.14	0.97	0.75	0.68
AT 14 %		1.12	0.96	0.75	0.67
AT 16 %		1.09	0.94	0.74	0.66
AT 18 %		1.07	0.93	0.73	0.66

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS		REVENUES			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Gold Metal		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-INF	-INF	8.76	26.74
TERMINAL WORTH	$	36	36	36	36
PAYBACK PERIOD	YEARS	16.00	16.00	5.89	1.60
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.65	0.78	1.06	1.20
AT 2 %		0.64	0.77	1.05	1.18
AT 4 %		0.63	0.76	1.03	1.17
AT 6 %		0.62	0.75	1.02	1.15
AT 8 %		0.61	0.74	1.01	1.14
AT 10 %		0.60	0.73	0.99	1.12
AT 12 %		0.59	0.72	0.98	1.10
AT 14 %		0.59	0.71	0.96	1.09
AT 16 %		0.58	0.70	0.95	1.07
AT 18 %		0.57	0.69	0.94	1.06

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Custom Mill - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS CAPITAL COST

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-18.76	-20.05	-22.01	-22.78
TERMINAL WORTH	$	25	31	41	47
PAYBACK PERIOD	YEARS	16.00	16.00	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.94	0.93	0.91	0.90
AT 2 %		0.94	0.92	0.90	0.89
AT 4 %		0.93	0.91	0.89	0.87
AT 6 %		0.92	0.90	0.87	0.86
AT 8 %		0.91	0.89	0.86	0.84
AT 10 %		0.90	0.88	0.85	0.83
AT 12 %		0.89	0.87	0.83	0.81
AT 14 %		0.88	0.86	0.82	0.80
AT 16 %		0.87	0.85	0.81	0.79
AT 18 %		0.86	0.84	0.79	0.77

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DESCRIPTION :
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                              Refractory Gold Processing Options
                       Economic Analysis - Constant CDN Dollars ('000, 1989)
                      Integrated Plant - Acidic Pressure Oxidation Technology
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
ANNUAL EXPENSES											
Ore Purchase		15521	15521	15521	15521	15521	15521	15521	15521	15521	15521
Labour		5084	5084	5084	5084	5084	5084	5084	5084	5084	5084
Reagents		7444	7444	7444	7444	7444	7444	7444	7444	7444	7444
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		910	910	910	910	910	910	910	910	910	910
Supplies/Insurance		2470	2470	2470	2470	2470	2470	2470	2470	2470	2470
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		32148	32148	32148	32148	32148	32148	32148	32148	32148	32148
CAPITAL OUTLAYS											
Purchase of Equipment		19379	0	0	0	0	0	0	0	0	0
Installation cost		10125	0	0	0	0	0	0	0	0	0
Instrumentation		2907	0	0	0	0	0	0	0	0	0
Piping		3876	0	0	0	0	0	0	0	0	0
Electrical		3876	0	0	0	0	0	0	0	0	0
Plant Services		1938	0	0	0	0	0	0	0	0	0
Building		1150	0	0	0	0	0	0	0	0	0
Land		120	0	0	0	0	0	0	0	0	0
Engineering		8674	0	0	0	0	0	0	0	0	0
Construction indirects		1301	0	0	0	0	0	0	0	0	0
Commissioning/Spares		867	0	0	0	0	0	0	0	0	0
Contingency		4337	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		5855	0	0	0	0	0	0	0	0	0
SUBTOTAL		64405	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		96553	32148	32148	32148	32148	32148	32148	32148	32148	32148
CASH FLOW		-60553	3852	3852	3852	3852	3852	3852	3852	3852	3852
ACCUMULATED CASH FLOW		-60553	-56701	-52849	-48997	-45145	-41293	-37441	-33589	-29737	-25885

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
TOTAL		36000	36000	36000	36000	36000
ANNUAL EXPENSES						
Ore Purchase		15521	15521	15521	15521	15521
Labour		5084	5084	5084	5084	5084
Reagents		7444	7444	7444	7444	7444
Acid Plant		0	0	0	0	0
Electricity		910	910	910	910	910
Supplies/Insurance		2470	2470	2470	2470	2470
Refining Charge		719	719	719	719	719
SUBTOTAL		32148	32148	32148	32148	32148
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0			
				0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		32148	32148	32148	32148	32148
CASH FLOW		3852	3852	3852	3852	3852
ACCUMULATED CASH FLOW		-22033	-18181	-14329	-10477	-6625

ECONOMIC RATE OF RETURN	-1.49
PAYBACK PERIOD	> 16.00
TERMINAL WORTH	60

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
TOTAL		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
ANNUAL EXPENSES											
Ore Purchase		232815	203422	179471	159789	143480	129859	118397	108679	100383	93251
Labour		76260	66632	58787	52340	46998	42536	38782	35599	32881	30545
Reagents		111660	97563	86076	76636	68814	62282	56784	52123	48144	44724
Acid Plant		0	0	0	0	0	0	0	0	0	0
Electricity		13650	11927	10522	9368	8412	7614	6942	6372	5895	5467
Supplies/Insurance		37050	32372	28561	25429	22833	20666	18842	17295	15975	14840
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		482220	421340	371731	330963	297184	268972	245230	225103	207918	193147
CAPITAL OUTLAYS											
Purchase of Equipment		19379	19379	19379	19379	19379	19379	19379	19379	19379	19379
Installation cost		10125	10125	10125	10125	10125	10125	10125	10125	10125	10125
Instrumentation		2907	2907	2907	2907	2907	2907	2907	2907	2907	2907
Piping		3876	3876	3876	3876	3876	3876	3876	3876	3876	3876
Electrical		3876	3876	3876	3876	3876	3876	3876	3876	3876	3876
Plant Services		1938	1938	1938	1938	1938	1938	1938	1938	1938	1938
Building		1150	1150	1150	1150	1150	1150	1150	1150	1150	1150
Land		120	120	120	120	120	120	120	120	120	120
Engineering		8674	8674	8674	8674	8674	8674	8674	8674	8674	8674
Construction indirects		1301	1301	1301	1301	1301	1301	1301	1301	1301	1301
Commissioning/Spares		867	867	867	867	867	867	867	867	867	867
Contingency		4337	4337	4337	4337	4337	4337	4337	4337	4337	4337
WORKING CAPITAL		5855	5855	5855	5855	5855	5855	5855	5855	5855	5855
TERMINAL WORTH		-60	-45	-35	-27	-20	-16	-12	-10	-8	-6
SUBTOTAL		64345	64360	64370	64378	64385	64389	64393	64395	64397	64399
TOTAL ANNUAL COSTS		546565	485699	436102	395342	361568	333361	309623	289498	272316	257546
NET PRESENT WORTH:		-6565	-13874	-19829	-24722	-28776	-32161	-35009	-37423	-39485	-41256
BENEFIT COST RATIO		0.99	0.97	0.95	0.94	0.92	0.90	0.89	0.87	0.86	0.84

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS		OPERATING COSTS			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Ore Purchase		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	25.39	12.61	-INF	-INF
TERMINAL WORTH	$	60	60	60	60
PAYBACK PERIOD	YEARS	1.77	4.42	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.34	1.14	0.87	0.78
AT 2 %		1.31	1.12	0.86	0.77
AT 4 %		1.28	1.09	0.85	0.76
AT 6 %		1.25	1.07	0.83	0.75
AT 8 %		1.22	1.05	0.82	0.74
AT 10 %		1.19	1.03	0.81	0.73
AT 12 %		1.16	1.01	0.79	0.72
AT 14 %		1.14	0.99	0.78	0.71
AT 16 %		1.11	0.97	0.77	0.70
AT 18 %		1.08	0.95	0.75	0.69

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS — REVENUES

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Gold Metal		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-INF	-INF	14.14	28.57
TERMINAL WORTH	$	60	60	60	60
PAYBACK PERIOD	YEARS	16.00	16.00	3.96	1.40
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.69	0.84	1.14	1.28
AT 2 %		0.68	0.83	1.12	1.26
AT 4 %		0.67	0.81	1.10	1.24
AT 6 %		0.66	0.80	1.08	1.22
AT 8 %		0.64	0.78	1.06	1.20
AT 10 %		0.63	0.77	1.04	1.17
AT 12 %		0.62	0.75	1.02	1.15
AT 14 %		0.61	0.74	1.00	1.13
AT 16 %		0.60	0.73	0.98	1.11
AT 18 %		0.59	0.71	0.97	1.09

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Acidic Pressure Oxidation Technology

SENSITIVITY ANALYSIS

CAPITAL COST

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	3.18	0.56	-3.16	-4.57
TERMINAL WORTH	$	42	51	69	78
PAYBACK PERIOD	YEARS	9.16	11.43	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.02	1.00	0.97	0.96
AT 2 %		1.01	0.99	0.95	0.94
AT 4 %		0.99	0.97	0.94	0.92
AT 6 %		0.98	0.96	0.92	0.90
AT 8 %		0.97	0.94	0.90	0.88
AT 10 %		0.95	0.93	0.88	0.86
AT 12 %		0.94	0.91	0.86	0.84
AT 14 %		0.93	0.90	0.85	0.82
AT 16 %		0.91	0.88	0.83	0.80
AT 18 %		0.90	0.87	0.81	0.79

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                    ***          OUTPUT LISTING            ***
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         TIME_____:18:16:36

DESCRIPTION :
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                                        Refractory Gold Processing Options
                                  Economic Analysis - Constant CDN Dollars ('000, 1989)
                                        Custom Mill - Fluidized Bed Roasting
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
Sulfuric Acid		2457	2457	2457	2457	2457	2457	2457	2457	2457	2457
TOTAL		38457	38457	38457	38457	38457	38457	38457	38457	38457	38457
ANNUAL EXPENSES											
Concentrate		27008	27008	27008	27008	27008	27008	27008	27008	27008	27008
Labour		3633	3633	3633	3633	3633	3633	3633	3633	3633	3633
Reagents		1289	1289	1289	1289	1289	1289	1289	1289	1289	1289
Acid Plant		1970	1970	1970	1970	1970	1970	1970	1970	1970	1970
Electricity		476	476	476	476	476	476	476	476	476	476
Supplies/Insurance		1761	1761	1761	1761	1761	1761	1761	1761	1761	1761
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		36856	36856	36856	36856	36856	36856	36856	36856	36856	36856
CAPITAL OUTLAYS											
Purchase of Equipment		13383	0	0	0	0	0	0	0	0	0
Installation cost		10118	0	0	0	0	0	0	0	0	0
Instrumentation		2007	0	0	0	0	0	0	0	0	0
Piping		1338	0	0	0	0	0	0	0	0	0
Electrical		2007	0	0	0	0	0	0	0	0	0
Plant Services		1338	0	0	0	0	0	0	0	0	0
Building		690	0	0	0	0	0	0	0	0	0
Land		14	0	0	0	0	0	0	0	0	0
Engineering		6179	0	0	0	0	0	0	0	0	0
Construction indirects		927	0	0	0	0	0	0	0	0	0
Commissioning/Spares		618	0	0	0	0	0	0	0	0	0
Contingency		4634	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		4325	0	0	0	0	0	0	0	0	0
SUBTOTAL		47578	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		84434	36856	36856	36856	36856	36856	36856	36856	36856	36856
CASH FLOW		-45977	1601	1601	1601	1601	1601	1601	1601	1601	1601
ACCUMULATED CASH FLOW		-45977	-44376	-42775	-41174	-39573	-37972	-36371	-34770	-33169	-31568

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
Sulfuric Acid		2457	2457	2457	2457	2457
TOTAL		38457	38457	38457	38457	38457
ANNUAL EXPENSES						
Concentrate		27008	27008	27008	27008	27008
Labour		3633	3633	3633	3633	3633
Reagents		1289	1289	1289	1289	1289
Acid Plant		1970	1970	1970	1970	1970
Electricity		476	476	476	476	476
Supplies/Insurance		1761	1761	1761	1761	1761
Refining Charge		719	719	719	719	719
SUBTOTAL		36856	36856	36856	36856	36856
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0	0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		36856	36856	36856	36856	36856
CASH FLOW		1601	1601	1601	1601	1601
ACCUMULATED CASH FLOW		-29967	-28366	-26765	-25164	-23563

ECONOMIC RATE OF RETURN	-6.38
PAYBACK PERIOD	> 16.00
TERMINAL WORTH	7

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
Sulfuric Acid		36855	32202	28411	25295	22713	20557	18742	17204	15891	14762
TOTAL		576855	504027	444683	395914	355506	321758	293357	269279	248722	231052
ANNUAL EXPENSES											
Concentrate		405120	353974	312297	278047	249668	225968	206022	189112	174675	162266
Labour		54495	47615	42009	37402	33584	30396	27713	25439	23497	21827
Reagents		19335	16894	14905	13270	11916	10785	9833	9026	8337	7744
Acid Plant		29550	25819	22779	20281	18211	16482	15027	13794	12741	11836
Electricity		7140	6239	5504	4900	4400	3983	3631	3333	3079	2860
Supplies/Insurance		26415	23080	20363	18129	16279	14734	13433	12331	11389	10580
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		552840	483044	426171	379432	340706	308363	281144	258068	238367	221433
CAPITAL OUTLAYS											
Purchase of Equipment		13383	13383	13383	13383	13383	13383	13383	13383	13383	13383
Installation cost		10118	10118	10118	10118	10118	10118	10118	10118	10118	10118
Instrumentation		2007	2007	2007	2007	2007	2007	2007	2007	2007	2007
Piping		1338	1338	1338	1338	1338	1338	1338	1338	1338	1338
Electrical		2007	2007	2007	2007	2007	2007	2007	2007	2007	2007
Plant Services		1338	1338	1338	1338	1338	1338	1338	1338	1338	1338
Building		690	690	690	690	690	690	690	690	690	690
Land		14	14	14	14	14	14	14	14	14	14
Engineering		6179	6179	6179	6179	6179	6179	6179	6179	6179	6179
Construction indirects		927	927	927	927	927	927	927	927	927	927
Commissioning/Spares		618	618	618	618	618	618	618	618	618	618
Contingency		4634	4634	4634	4634	4634	4634	4634	4634	4634	4634
WORKING CAPITAL		4325	4325	4325	4325	4325	4325	4325	4325	4325	4325
TERMINAL WORTH		-7	-5	-4	-3	-2	-2	-1	-1	-1	-1
SUBTOTAL		47571	47573	47574	47575	47576	47576	47577	47577	47577	47577
TOTAL ANNUAL COSTS		600411	530617	473745	427007	388281	355939	328720	305645	285944	269010
NET PRESENT WORTH:		-23556	-26590	-29061	-31093	-32776	-34181	-35364	-36367	-37223	-37958
BENEFIT COST RATIO		0.96	0.95	0.94	0.93	0.92	0.90	0.89	0.88	0.87	0.86

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

SENSITIVITY ANALYSIS — OPERATING COSTS

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Concentrate		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	35.74	15.19	-INF	-INF
TERMINAL WORTH	$	7	7	7	7
PAYBACK PERIOD	YEARS	0.76	3.67	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.33	1.11	0.84	0.75
AT 2 %		1.31	1.10	0.84	0.75
AT 4 %		1.29	1.09	0.83	0.74
AT 6 %		1.26	1.07	0.82	0.73
AT 8 %		1.24	1.05	0.81	0.72
AT 10 %		1.22	1.04	0.80	0.72
AT 12 %		1.20	1.02	0.79	0.71
AT 14 %		1.18	1.01	0.78	0.70
AT 16 %		1.16	0.99	0.77	0.70
AT 18 %		1.14	0.98	0.76	0.69

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

SENSITIVITY ANALYSIS		REVENUES			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Gold Metal		-30.00	-15.00	15.00	30.00
Sulfuric Acid		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-INF	-INF	16.04	37.71
TERMINAL WORTH	$	7	7	7	7
PAYBACK PERIOD	YEARS	16.00	16.00	3.46	0.62
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.67	0.82	1.10	1.25
AT 2 %		0.66	0.81	1.09	1.23
AT 4 %		0.66	0.80	1.08	1.22
AT 6 %		0.65	0.79	1.07	1.21
AT 8 %		0.64	0.78	1.05	1.19
AT 10 %		0.63	0.77	1.04	1.18
AT 12 %		0.62	0.76	1.03	1.16
AT 14 %		0.62	0.75	1.01	1.15
AT 16 %		0.61	0.74	1.00	1.13
AT 18 %		0.60	0.73	0.99	1.12

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Fluidized Bed Roasting Technology

SENSITIVITY ANALYSIS		CAPITAL COST			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	-4.77	-6.77	-9.71	-10.84
TERMINAL WORTH	$	5	6	8	9
PAYBACK PERIOD	YEARS	16.00	16.00	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		0.98	0.97	0.95	0.94
AT 2 %		0.97	0.96	0.94	0.93
AT 4 %		0.97	0.95	0.93	0.91
AT 6 %		0.96	0.94	0.91	0.90
AT 8 %		0.95	0.93	0.90	0.89
AT 10 %		0.94	0.92	0.89	0.87
AT 12 %		0.93	0.91	0.88	0.86
AT 14 %		0.92	0.90	0.86	0.85
AT 16 %		0.91	0.89	0.85	0.83
AT 18 %		0.90	0.88	0.84	0.82

```
                              ******************************************
                              ***            OUTPUT LISTING           ***
                              ******************************************




                              ******************************************
                              ***  K A S H P A K    P R O G R A M    ***
                              ******************************************



         FILE NAME :intfbd

         DATE_____:03-29-1989

         TIME_____:17:24:56

DESCRIPTION :
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                                        Refractory Gold Processing Options
                                   Economic Analysis - Constant CDN Dollars ('000, 1989)
                                   Integrated Plant - Fluidized Bed Roasting Technology
---------------------------------------------------------------------------------------------------------------
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Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Fluidized Bed Roasting Technology

ECONOMIC ANALYSIS

	L/F	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998
ANNUAL BENEFITS											
Gold Metal		36000	36000	36000	36000	36000	36000	36000	36000	36000	36000
Sulfuric Acid		2457	2457	2457	2457	2457	2457	2457	2457	2457	2457
TOTAL		38457	38457	38457	38457	38457	38457	38457	38457	38457	38457
ANNUAL EXPENSES											
Ore Purchase		19676	19676	19676	19676	19676	19676	19676	19676	19676	19676
Labour		5084	5084	5084	5084	5084	5084	5084	5084	5084	5084
Reagents		1852	1852	1852	1852	1852	1852	1852	1852	1852	1852
Acid Plant		1970	1970	1970	1970	1970	1970	1970	1970	1970	1970
Electricity		1084	1084	1084	1084	1084	1084	1084	1084	1084	1084
Supplies/Insurance		2408	2408	2408	2408	2408	2408	2408	2408	2408	2408
Refining Charge		719	719	719	719	719	719	719	719	719	719
SUBTOTAL		32793	32793	32793	32793	32793	32793	32793	32793	32793	32793
CAPITAL OUTLAYS											
Purchase of Equipment		18524	0	0	0	0	0	0	0	0	0
Installation cost		12586	0	0	0	0	0	0	0	0	0
Instrumentation		2779	0	0	0	0	0	0	0	0	0
Piping		1852	0	0	0	0	0	0	0	0	0
Electrical		2779	0	0	0	0	0	0	0	0	0
Plant Services		1852	0	0	0	0	0	0	0	0	0
Building		1150	0	0	0	0	0	0	0	0	0
Land		24	0	0	0	0	0	0	0	0	0
Engineering		8309	0	0	0	0	0	0	0	0	0
Construction indirects		1246	0	0	0	0	0	0	0	0	0
Commissioning/Spares		831	0	0	0	0	0	0	0	0	0
Contingency		4155	0	0	0	0	0	0	0	0	0
WORKING CAPITAL		5609	0	0	0	0	0	0	0	0	0
SUBTOTAL		61696	0	0	0	0	0	0	0	0	0
TOTAL ANNUAL COSTS		94489	32793	32793	32793	32793	32793	32793	32793	32793	32793
CASH FLOW		-56032	5664	5664	5664	5664	5664	5664	5664	5664	5664
ACCUMULATED CASH FLOW		-56032	-50368	-44704	-39040	-33376	-27712	-22048	-16384	-10720	-5056

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Fluidized Bed Roasting Technology

ECONOMIC ANALYSIS

	L/F	1999	2000	2001	2002	2003
ANNUAL BENEFITS						
Gold Metal		36000	36000	36000	36000	36000
Sulfuric Acid		2457	2457	2457	2457	2457
TOTAL		38457	38457	38457	38457	38457
ANNUAL EXPENSES						
Ore Purchase		19676	19676	19676	19676	19676
Labour		5084	5084	5084	5084	5084
Reagents		1852	1852	1852	1852	1852
Acid Plant		1970	1970	1970	1970	1970
Electricity		1084	1084	1084	1084	1084
Supplies/Insurance		2408	2408	2408	2408	2408
Refining Charge		719	719	719	719	719
SUBTOTAL		32793	32793	32793	32793	32793
CAPITAL OUTLAYS						
Purchase of Equipment		0	0	0	0	0
Installation cost		0	0	0	0	0
Instrumentation		0	0	0	0	0
Piping		0	0	0	0	0
Electrical		0	0	0	0	0
Plant Services		0	0	0	0	0
Building		0	0	0	0	0
Land		0	0	0	0	0
Engineering		0	0	0	0	0
Construction indirects		0	0	0	0	0
Commissioning/Spares		0	0	0	0	0
Contingency		0	0	0	0	0
WORKING CAPITAL		0	0	0	0	0
SUBTOTAL		0	0	0	0	0
TOTAL ANNUAL COSTS		32793	32793	32793	32793	32793
CASH FLOW		5664	5664	5664	5664	5664
ACCUMULATED CASH FLOW		608	6272	11936	17600	23264

ECONOMIC RATE OF RETURN	5.01
PAYBACK PERIOD	7.99
TERMINAL WORTH	12

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Fluidized Bed Roasting Technology

TOTAL DISCOUNTED VALUES FOR SELECTED RATES

DISCOUNT RATES	L/F	0 %	2 %	4 %	6 %	8 %	10 %	12 %	14 %	16 %	18 %
ANNUAL BENEFITS											
Gold Metal		540000	471825	416273	370620	332793	301201	274614	252075	232831	216290
Sulfuric Acid		36855	32202	28411	25295	22713	20557	18742	17204	15891	14762
TOTAL		576855	504027	444683	395914	355506	321758	293357	269279	248722	231052
ANNUAL EXPENSES											
Ore Purchase		295140	257879	227516	202564	181890	164623	150092	137773	127255	118215
Labour		76260	66632	58787	52340	46999	42535	38792	35599	32881	30545
Reagents		27780	24273	21415	19066	17120	15495	14127	12968	11978	11127
Acid Plant		29550	25819	22779	20231	18211	16492	15027	13794	12741	11836
Electricity		16260	14207	12534	11160	10021	9069	8269	7590	7011	6513
Supplies/Insurance		36120	31560	27844	24790	22260	20147	18369	16861	15574	14467
Refining Charge		10785	9423	8314	7402	6647	6016	5485	5034	4650	4320
SUBTOTAL		491895	429793	379190	337604	303146	274369	250151	229619	212090	197022
CAPITAL OUTLAYS											
Purchase of Equipment		18524	18524	18524	18524	18524	18524	18524	18524	18524	18524
Installation cost		12586	12586	12586	12586	12586	12586	12586	12586	12586	12586
Instrumentation		2779	2779	2779	2779	2779	2779	2779	2779	2779	2779
Piping		1852	1852	1852	1852	1852	1852	1852	1852	1852	1852
Electrical		2779	2779	2779	2779	2779	2779	2779	2779	2779	2779
Plant Services		1852	1852	1852	1852	1852	1852	1852	1852	1852	1852
Building		1150	1150	1150	1150	1150	1150	1150	1150	1150	1150
Land		24	24	24	24	24	24	24	24	24	24
Engineering		8309	8309	8309	8309	8309	8309	8309	8309	8309	8309
Construction indirects		1246	1246	1246	1246	1246	1246	1246	1246	1246	1246
Commissioning/Spares		831	831	831	831	831	831	831	831	831	831
Contingency		4155	4155	4155	4155	4155	4155	4155	4155	4155	4155
WORKING CAPITAL		5609	5609	5609	5609	5609	5609	5609	5609	5609	5609
TERMINAL WORTH		-12	-9	-7	-5	-4	-3	-2	-2	-2	-1
SUBTOTAL		61684	61687	61689	61691	61692	61693	61694	61694	61695	61695
TOTAL ANNUAL COSTS		553579	491480	440879	399294	364838	336062	311844	291313	273784	258717
NET PRESENT WORTH:		23276	12547	3805	-3380	-9332	-14304	-18488	-22034	-25062	-27665
BENEFIT COST RATIO		1.04	1.03	1.01	0.99	0.97	0.96	0.94	0.92	0.91	0.89

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Fluidized Bed Roasting Technology

SENSITIVITY ANALYSIS		OPERATING COSTS			
ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Ore Purchase		-30.00	-15.00	15.00	30.00
Labour		-30.00	-15.00	15.00	30.00
Reagents		-30.00	-15.00	15.00	30.00
Acid Plant		-30.00	-15.00	15.00	30.00
Electricity		-30.00	-15.00	15.00	30.00
Supplies/Insurance		-30.00	-15.00	15.00	30.00
Refining Charge		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	32.93	18.96	-17.78	-INF
TERMINAL WORTH	$	12	12	12	12
PAYBACK PERIOD	YEARS	0.98	2.83	16.00	16.00
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.42	1.20	0.92	0.82
AT 2 %		1.39	1.18	0.91	0.81
AT 4 %		1.36	1.16	0.89	0.80
AT 6 %		1.33	1.14	0.88	0.79
AT 8 %		1.30	1.11	0.87	0.78
AT 10 %		1.27	1.09	0.85	0.77
AT 12 %		1.24	1.07	0.84	0.76
AT 14 %		1.21	1.05	0.83	0.75
AT 16 %		1.18	1.03	0.81	0.74
AT 18 %		1.16	1.01	0.80	0.73

```
--------------------------------------------------------------------------------------------------------
                               Refractory Gold Processing Options
                      Economic Analysis - Constant CDN Dollars ('000, 1989)
                        Integrated Plant - Fluidized Bed Roasting Technology
--------------------------------------------------------------------------------------------------------

SENSITIVITY ANALYSIS                   REVENUES
--------------------                   ------------------------------

ANALYSIS No.                              1        2        3        4
--------------------------------------------------------------------------------------------------------
DESCRIPTION                  L/F      % CHANGE % CHANGE % CHANGE % CHANGE
--------------------------------------------------------------------------------------------------------
Gold Metal                             -30.00   -15.00    15.00    30.00
Sulfuric Acid                          -30.00   -15.00    15.00    30.00


ECONOMIC RATE OF RETURN     %          -INF     -INF      21.20    38.24
TERMINAL WORTH              $              12       12       12       12
PAYBACK PERIOD              YEARS       16.00    16.00     2.40     0.59
DISCOUNT RATES
-------------
BENEFIT COST RATIO

   AT  0  %                              0.73     0.99     1.20     1.35
   AT  2  %                              0.72     0.87     1.18     1.33
   AT  4  %                              0.71     0.86     1.16     1.31
   AT  6  %                              0.69     0.84     1.14     1.29
   AT  8  %                              0.68     0.83     1.12     1.27
   AT  10  %                             0.67     0.81     1.10     1.24
   AT  12  %                             0.56     0.90     1.08     1.22
   AT  14  %                             0.65     0.79     1.06     1.20
   AT  16  %                             0.64     0.77     1.04     1.18
   AT  18  %                             0.63     0.76     1.03     1.15
```

Refractory Gold Processing Options
Economic Analysis - Constant CDN Dollars ('000, 1989)
Integrated Plant - Fluidized Bed Roasting Technology

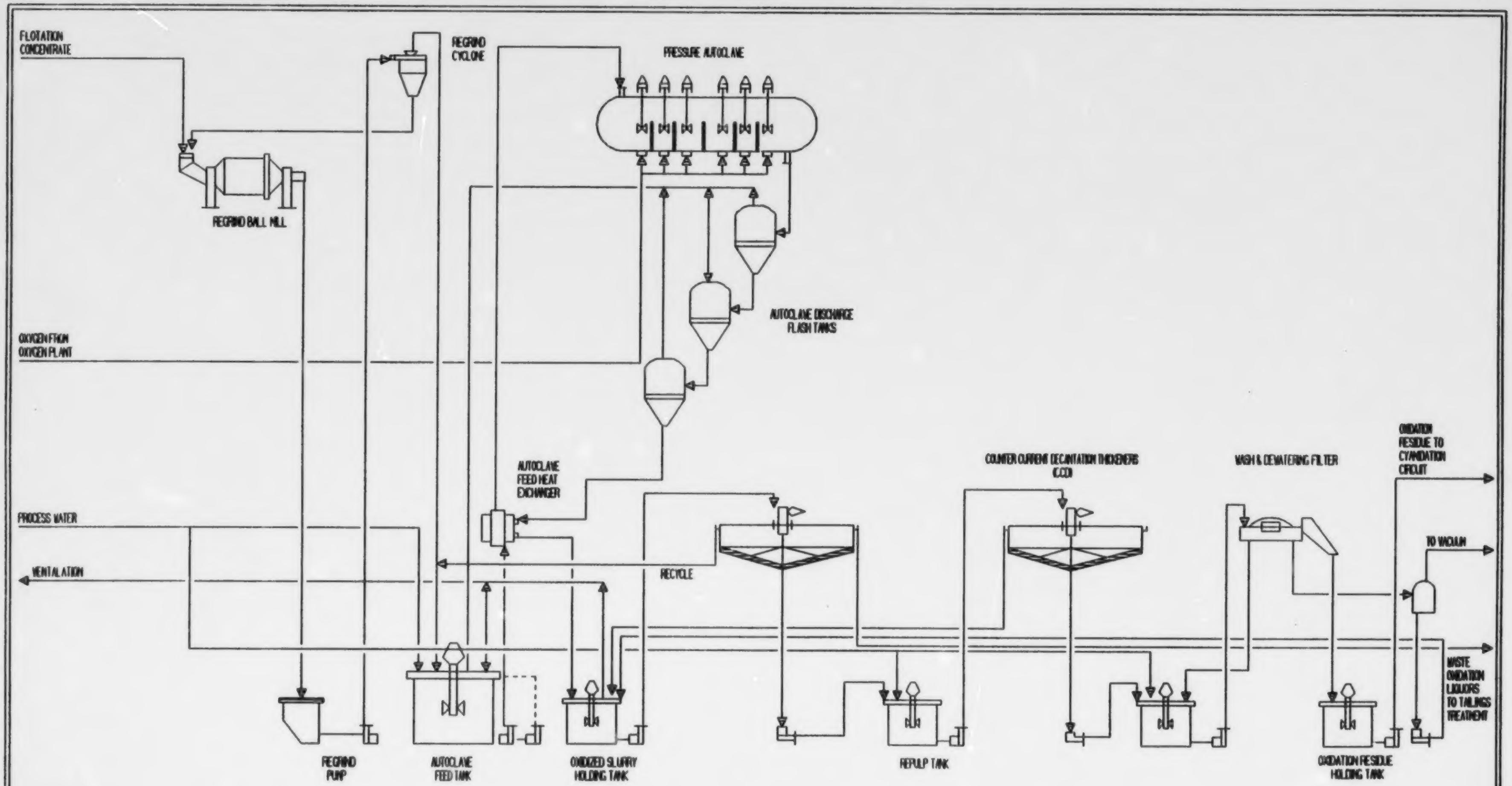
SENSITIVITY ANALYSIS — CAPITAL COST

ANALYSIS No.		1	2	3	4
DESCRIPTION	L/F	% CHANGE	% CHANGE	% CHANGE	% CHANGE
Purchase of Equipment		-30.00	-15.00	15.00	30.00
Installation cost		-30.00	-15.00	15.00	30.00
Instrumentation		-30.00	-15.00	15.00	30.00
Piping		-30.00	-15.00	15.00	30.00
Electrical		-30.00	-15.00	15.00	30.00
Plant Services		-30.00	-15.00	15.00	30.00
Building		-30.00	-15.00	15.00	30.00
Land		-30.00	-15.00	15.00	30.00
Engineering		-30.00	-15.00	15.00	30.00
Construction indirects		-30.00	-15.00	15.00	30.00
Commissioning/Spares		-30.00	-15.00	15.00	30.00
Contingency		-30.00	-15.00	15.00	30.00
ECONOMIC RATE OF RETURN	%	11.16	7.66	2.89	1.15
TERMINAL WORTH	$	8	10	14	16
PAYBACK PERIOD	YEARS	4.92	6.41	9.38	10.86
DISCOUNT RATES					
BENEFIT COST RATIO					
AT 0 %		1.07	1.06	1.03	1.01
AT 2 %		1.06	1.04	1.01	0.99
AT 4 %		1.05	1.03	0.99	0.97
AT 6 %		1.04	1.01	0.97	0.95
AT 8 %		1.02	1.00	0.95	0.93
AT 10 %		1.01	0.98	0.93	0.91
AT 12 %		0.99	0.97	0.92	0.89
AT 14 %		0.98	0.95	0.90	0.87
AT 16 %		0.97	0.94	0.88	0.85
AT 18 %		0.96	0.92	0.86	0.84

APPENDIX I

Drawings

Process Flow and Plant General Arrangement


FLOTATION CONCENTRATE
REGRIND CYCLONE
PRESSURE AUTOCLAVE
REGRIND BALL MILL
AUTOCLAVE DISCHARGE FLASH TANKS
OXYGEN FROM OXYGEN PLANT
AUTOCLAVE FEED HEAT EXCHANGER
COUNTER CURRENT DECANTATION THICKENERS (C.C.D)
WASH & DEWATERING FILTER
OXIDATION RESIDUE TO CYANIDATION CIRCUIT
PROCESS WATER
TO VACUUM
VENTILATION
RECYCLE
WASTE OXIDATION LIQUORS TO TAILINGS TREATMENT
REGRIND PUMP
AUTOCLAVE FEED TANK
OXIDIZED SLURRY HOLDING TANK
REPULP TANK
OXIDATION RESIDUE HOLDING TANK

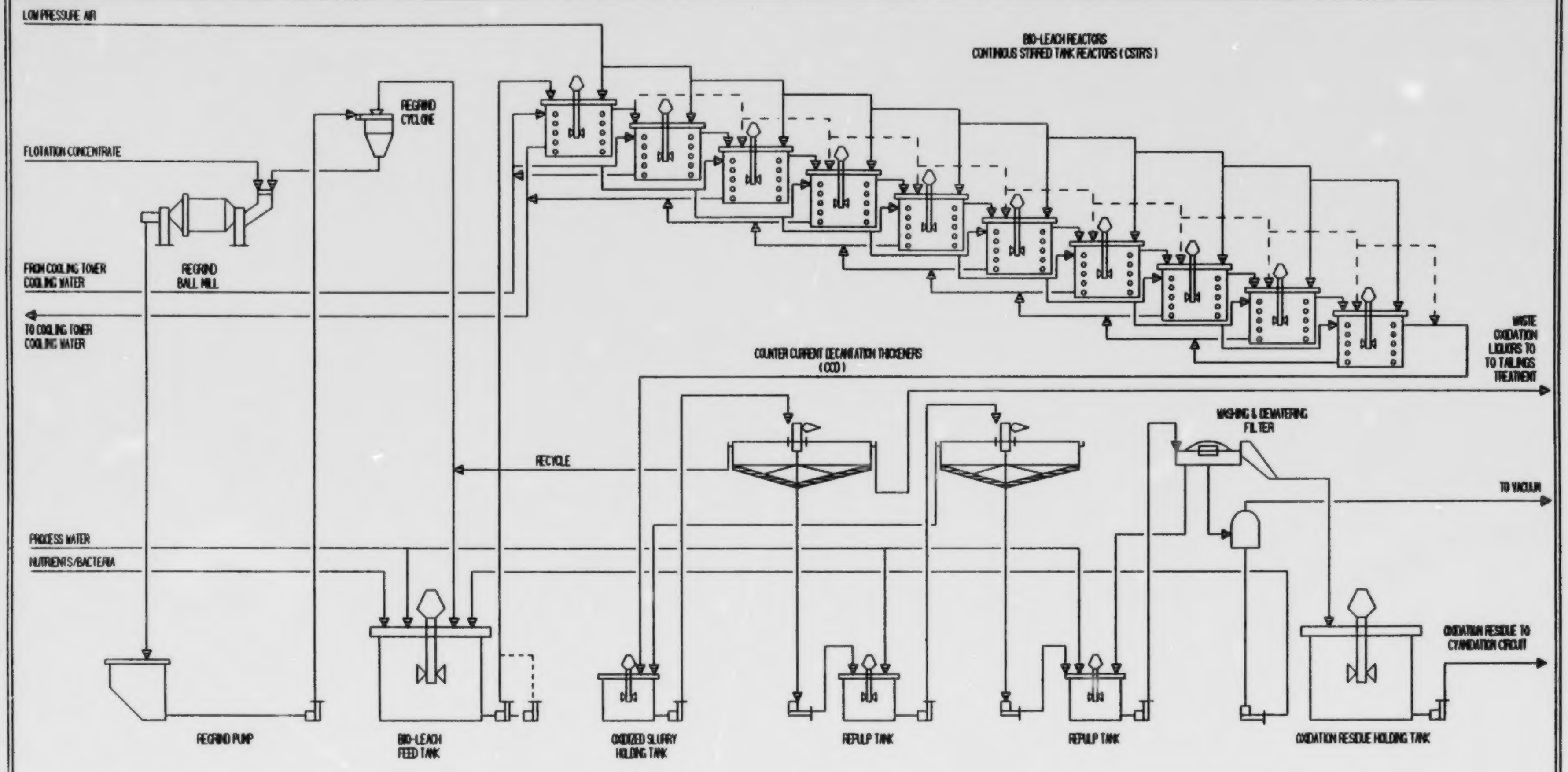
WHALEN THIBAULT
Consulting Engineers

Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

Drawing Name: Acidic Pressure Oxidation

Prepared by: TJW	Drawing No.: 01
Approved by: PJW JDT	Date: May 1989


LOW PRESSURE AIR
BIO-LEACH REACTORS
CONTINUOUS STIRRED TANK REACTORS (CSTR'S)
FLOTATION CONCENTRATE
REGRIND CYCLONE
REGRIND BALL MILL
FROM COOLING TOWER COOLING WATER
TO COOLING TOWER COOLING WATER
WASTE OXIDATION LIQUORS TO TO TAILINGS TREATMENT
COUNTER CURRENT DECANTATION THICKENERS (CCD)
WASHING & DEWATERING FILTER
RECYCLE
TO VACUUM
PROCESS WATER
NUTRIENTS/BACTERIA
OXIDATION RESIDUE TO CYANIDATION CIRCUIT
REGRIND PUMP
BIO-LEACH FEED TANK
OXIDIZED SLURRY HOLDING TANK
REPULP TANK
REPULP TANK
OXIDATION RESIDUE HOLDING TANK

WHALEN THIBAULT
Consulting Engineers

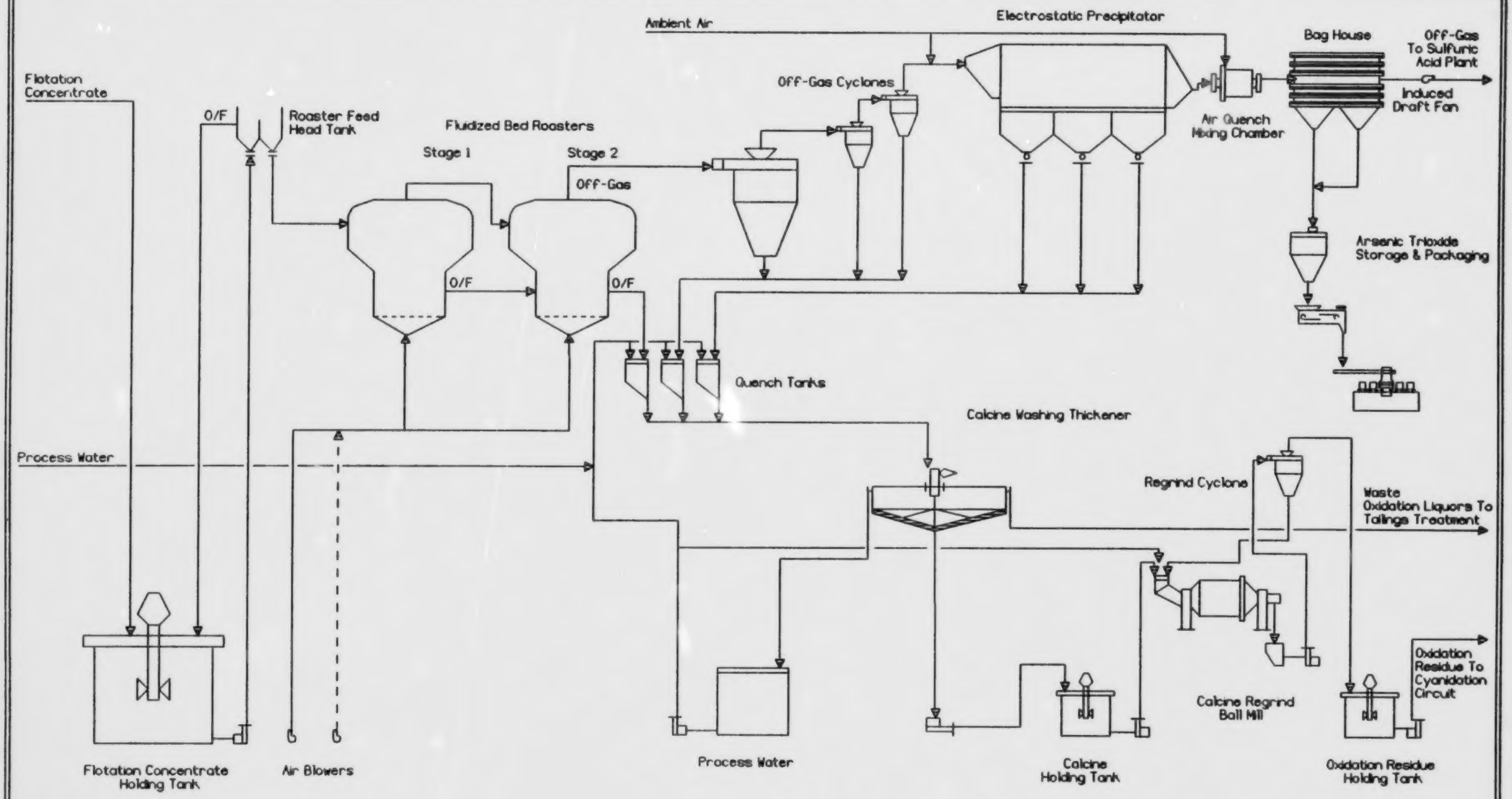
Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

Drawing Name: Biological Oxidation

Prepared by: TJW
Drawing No.: 02

Approved by: PJW JDT
Date: May 1989


Flotation Concentrate
O/F
Roaster Feed Head Tank
Fluidized Bed Roasters
Stage 1
Stage 2
Off-Gas
O/F
O/F
Ambient Air
Off-Gas Cyclones
Electrostatic Precipitator
Air Quench Mixing Chamber
Bag House
Off-Gas To Sulfuric Acid Plant
Induced Draft Fan
Arsenic Trioxide Storage & Packaging
Quench Tanks
Calcine Washing Thickener
Process Water
Regrind Cyclone
Waste Oxidation Liquors To Tailings Treatment
Oxidation Residue To Cyanidation Circuit
Calcine Regrind Ball Mill
Flotation Concentrate Holding Tank
Air Blowers
Process Water
Calcine Holding Tank
Oxidation Residue Holding Tank

WHALEN THIBAULT
Consulting Engineers

Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

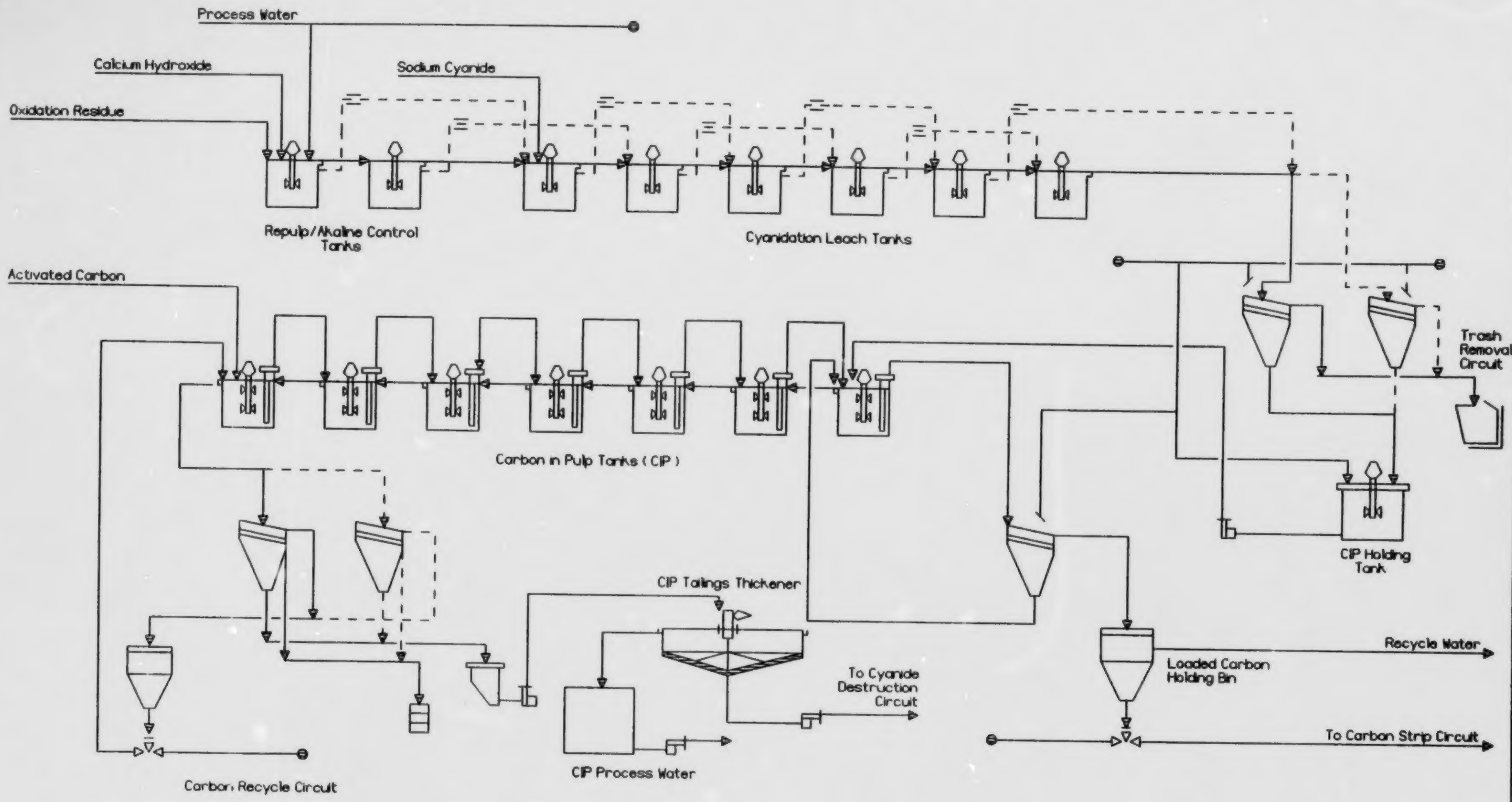
Drawing Name: Fluidized Bed Roasting

Prepared by: TJW

Drawing No.: 03

Approved by: PJW JDT

Date: May 1989


Process Water
Calcium Hydroxide
Oxidation Residue
Sodium Cyanide
Repulp/Alkaline Control Tanks
Cyanidation Leach Tanks
Activated Carbon
Trash Removal Circuit
Carbon in Pulp Tanks (CIP)
CIP Holding Tank
CIP Tailings Thickener
To Cyanide Destruction Circuit
Recycle Water
Loaded Carbon Holding Bin
To Carbon Strip Circuit
CIP Process Water
Carbon Recycle Circuit

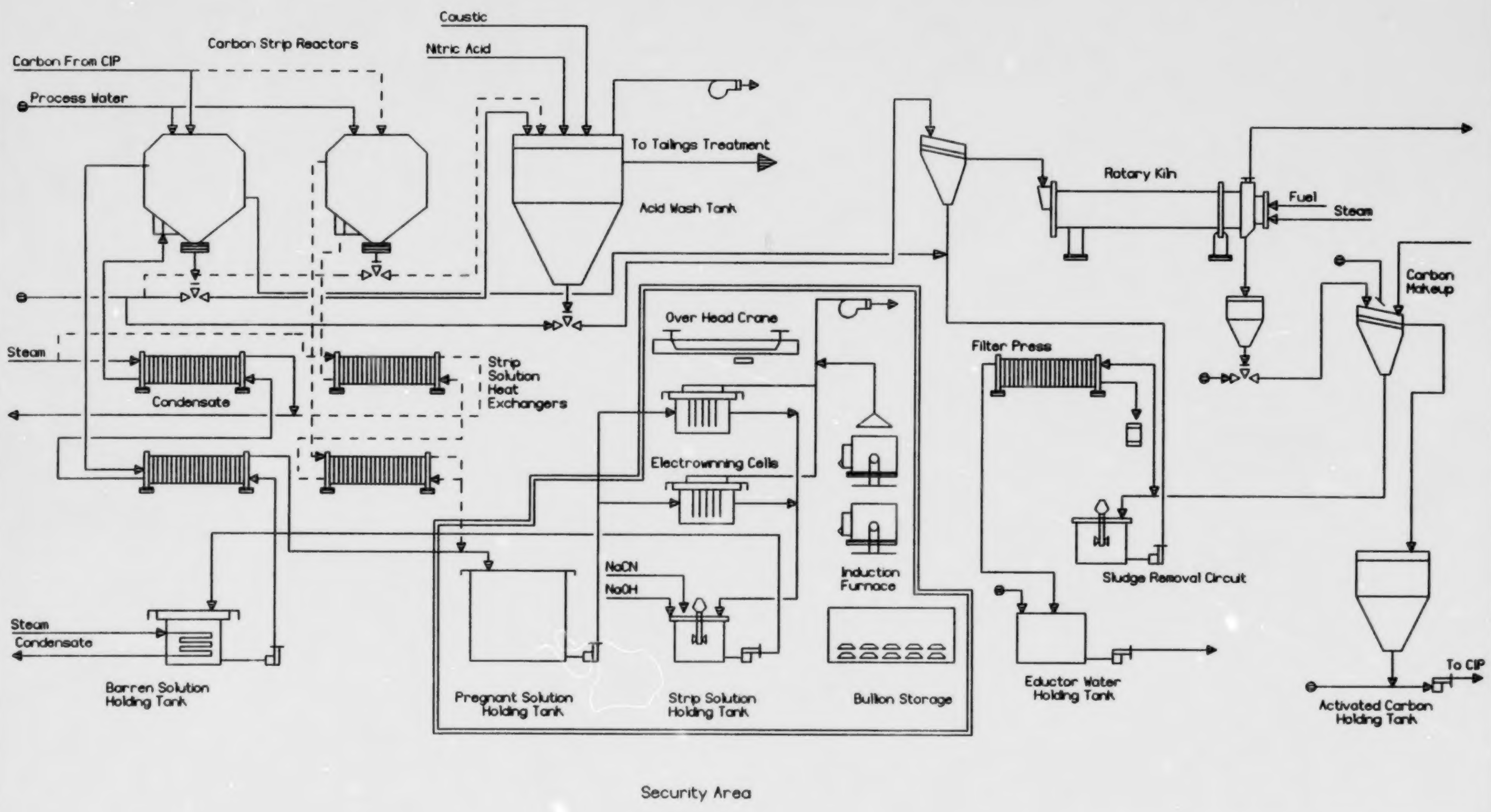
WHALEN THIBAULT
Consulting Engineers

Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

Drawing Name: Cyanidation Circuit

Prepared by: TJW	Drawing No.: 04
Approved by: PJW JDT	Date: May 1989


Carbon Strip Reactors
Carbon From CIP
Process Water
Caustic
Nitric Acid
To Tailings Treatment
Acid Wash Tank
Steam
Condensate
Strip Solution Heat Exchangers
Rotary Kiln
Fuel
Steam
Carbon Makeup
Over Head Crane
Electrowinning Cells
Induction Furnace
Filter Press
Sludge Removal Circuit
NaCN
NaOH
Steam
Condensate
Barren Solution Holding Tank
Pregnant Solution Holding Tank
Strip Solution Holding Tank
Bullion Storage
Eductor Water Holding Tank
Activated Carbon Holding Tank
To CIP
Security Area

WHALEN THIBAULT
Consulting Engineers

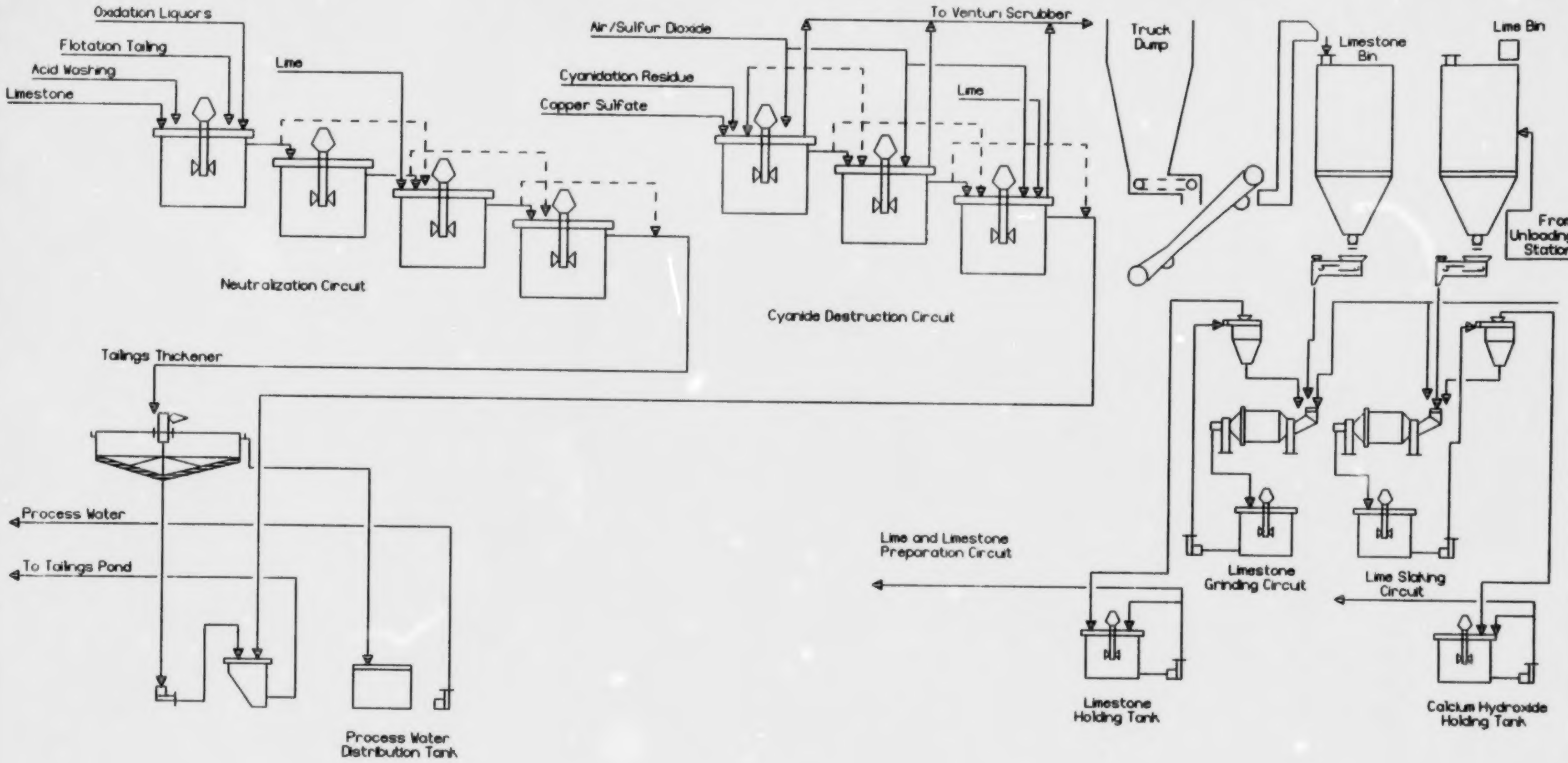
Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

Drawing Name: Carbon Treatment

Prepared by: TJW
Drawing No.: 05

Approved by: PJW JDT
Date: May 1989


Oxidation Liquors
Flotation Tailing
Acid Washing
Limestone
Lime
Neutralization Circuit
Air/Sulfur Dioxide
Cyanidation Residue
Copper Sulfate
To Venturi Scrubber
Lime
Cyanide Destruction Circuit
Truck Dump
Limestone Bin
Lime Bin
From Unloading Station
Tailings Thickener
Process Water
To Tailings Pond
Process Water Distribution Tank
Lime and Limestone Preparation Circuit
Limestone Grinding Circuit
Lime Slaking Circuit
Limestone Holding Tank
Calcium Hydroxide Holding Tank

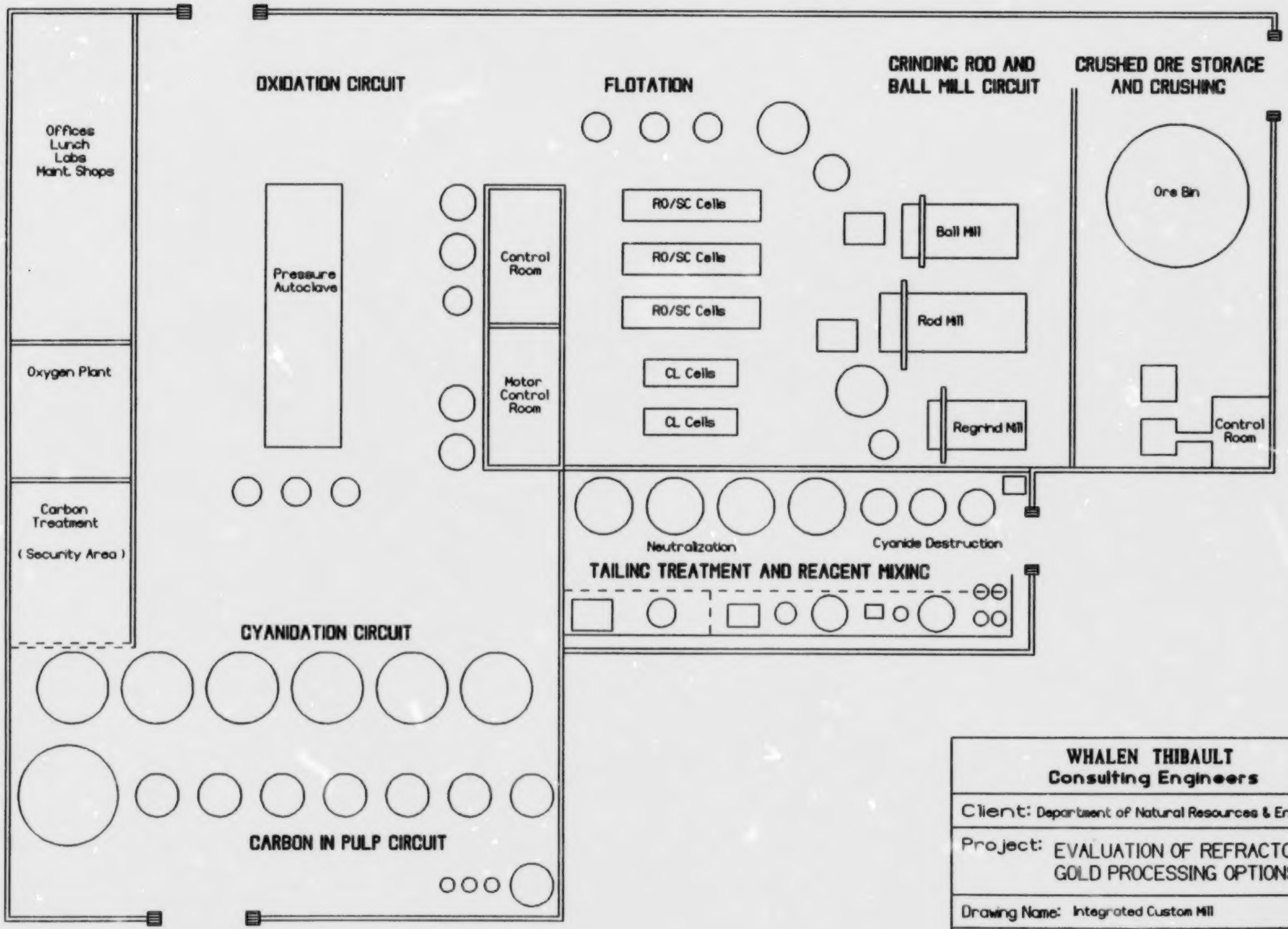
WHALEN THIBAULT
Consulting Engineers

Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY
GOLD PROCESSING OPTIONS

Drawing Name: Tailings Treatment

Prepared by: TJW
Drawing No.: 06
Approved by: PJW JDT
Date: May 1989


OXIDATION CIRCUIT
FLOTATION
GRINDING ROD AND BALL MILL CIRCUIT
CRUSHED ORE STORAGE AND CRUSHING
Offices
Lunch
Labs
Maint. Shops
Oxygen Plant
Carbon Treatment
(Security Area)
Pressure Autoclave
Control Room
Motor Control Room
RO/SC Cells
RO/SC Cells
RO/SC Cells
CL Cells
CL Cells
Ball Mill
Rod Mill
Regrind Mill
Ore Bin
Control Room
Neutralization
Cyanide Destruction
TAILING TREATMENT AND REAGENT MIXING
CYANIDATION CIRCUIT
CARBON IN PULP CIRCUIT

WHALEN THIBAULT
Consulting Engineers

Client: Department of Natural Resources & Energy

Project: EVALUATION OF REFRACTORY GOLD PROCESSING OPTIONS

Drawing Name: Integrated Custom Mill

Prepared by: TJW

Drawing No.: 07

Approved by: PJW JDT

Date: May 1989

VOLUME II

An Investigation of the Recovery of Gold from Elmtree Ore Samples

Prepared By:

Lakefield Research

An Investigation of

THE RECOVERY OF GOLD

from Elmtree ore samples

submitted by

WHALEN, THIBAULT & ASSOCIATES

Progress Report No. 1

Project No. L.R. 3608

NOTE:

This report refers to the samples as received.

LAKEFIELD RESEARCH
A DIVISION OF FALCONBRIDGE LIMITED
February 20th, 1989

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INTRODUCTION

This report contains the results of preliminary testwork conducted on Elmtree ore samples as requested by Mr. D. Thibault of Whalen, Thibault & Associates.

The purpose of the testwork was to investigate preconcentration of gold into a sulphide flotation concentrate and to further evaluate refractory gold extraction from the flotation concentrate by the following methods:

- cyanidation
- pressure oxidation
- pressure oxidation with nitric acid
- roasting
- biological leaching

Environmental studies on the arsenic stability compound were conducted for the pressure oxidation products.

The results were discussed with Mr. Thibault as they became available.

LAKEFIELD RESEARCH

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SUMMARY

1. Head Analysis

A representative sample was riffled out and analyzed for the following elements:

Gold, g/t Au	-	5.71 ± 0.22
Silver, g/t Ag	-	<2.0
Arsenic, % As	-	1.73
Sulphur, % S	-	4.02
Iron, % Fe	-	10.2

A semi-quantitative spectrographic analysis was conducted. The results were as follows:

Semi-Quantitative Spectrographic Analysis

10	-	100%	Si
3	-	30%	Fe
1	-	10%	Mg, Na
0.3	-	3%	Al, Ca
0.1	-	1%	As, Ti
0.03	-	0.3%	Mn
0.01	-	0.1%	K, Y
0.003	-	0.03%	Ga, Zr, Sr
0.001	-	0.01%	Tl, Pb, V, Cu, Cr
0.0003	-	0.003%	
0.0001	-	0.001%	
< 0.0003%			Ag, Ni, Co
I			
S			

I = Interference prevents positive identification

S = Strong spectral lines, unable to estimate amount

Unless specified above, the following were not detected at the approximate ppm lower limits of 0.5 Cu, Ag; 1 Mn; 5 Mg, Cr; 10 Be, Bi, Ca, Co, Ni, V; 25 Ge, Fe, Pb, Mo, Si, Sr, Sn, Ti, Zr, Tl, Pd, U, Th; 50 Al, Sb, Bi, Cd, Ga, Li, Zn; 100 As, Au, Ba, In, Na; 200 Nb, Ta, W, Rb, Pt; 300 P, Te, Y, Ce, 1000 K.

2. Mineralogy

A head sample was submitted for a brief mineralogical examination. The major opaque minerals and their associations were as follows:

Arsenopyrite occurred in fine to coarse grained euhedral crystals. Alteration to realgar or orpinent was not present.

Pyrrhotite was about 95% replaced by marcasite.

Goethite appeared to be a secondary replacement of pyrrhotite/marcasite and arsenopyrite.

Two small grains of chalcopyrite were present in the section.

No pyrite or native gold was observed.

Ilmenite was abundant, occurring as skeletal grains and irregular patches.

The degree of oxidation could not be quantitatively measured.

The sample was classified as refractory arsenopyrite ore.

3. Laboratory Testwork

The purpose of the laboratory testwork was to investigate preconcentration of gold into a sulphide flotation concentrate and to further evaluate refractory gold extraction from the flotation concentrate by the following methods:

1. Cyanidation - without regrind
   - with ultra-fine regrind
2. Pressure oxidation
3. Pressure oxidation with Nitric Acid
4. Roasting
5. Biological Leaching

Summary

It was beyond the scope of this report to select and to recommend one of the favourable methods. In the selection process the following factors should be considered:

- metallurgical response
- ore tonnage
- capital cost
- operating cost
- environmental treatment
- technical risk

3.1 Flotation

The recovery of gold into a sulphide concentrate was investigated by flotation . PAX and Aerofloat-208 were applied as collectors and MIBC was used as a frother. To improve gold recovery and to achieve the target specifications for concentrate grade in a range of 25 g/t Au, the effects of $CuSO_4$, Na_2CO_3, M2030 and NaHS additions were investigated.

Three small scale flotation tests on 2 kg charges were conducted to establish conditions and three tests on 10 kg samples were performed to produce flotation concentrate for further metallurgical testwork. The test conditions and results are summarized in Tables No. 1 and 2. To improve gold recovery, the rougher tailings were reground and floated again with PAX and a large dose of NaHS (1 kg/t) added as a sulphidizing agent. Overall gold recovery was improved by 2%; however, the concentrate grade decreased from 25 g/t to 22 g/t Au.

The final rougher tailing of Test 2 was submitted for mineralogical examination, which indicated that the pyrrhotite was completely oxidized to secondary products. This phenomenon suggested that the gold associated with pyrrhotite may restrict sulphide flotation recovery.

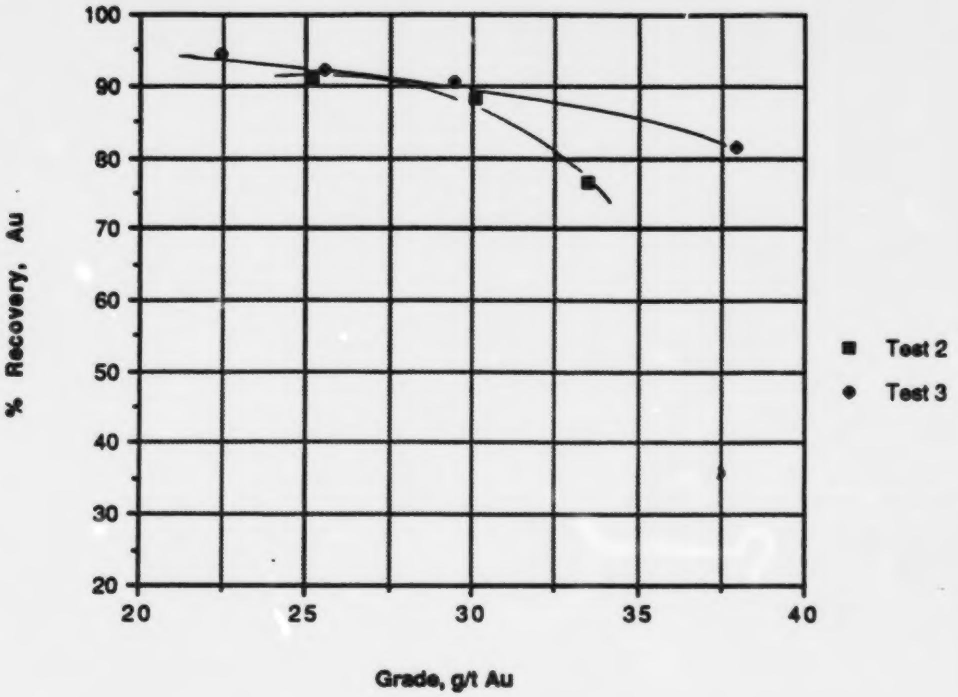
Table No. 1 - Flotation Test Conditions

Test No.	% -200 mesh	pH	Reagent Additions, g/t	
1	65.0	6.7	45 g/t AX-350;	45 g/t R-208, 20 g/t MIBC
2	86.4	6.5	80 g/t A-350; 250 g/t $CuSO_4$;	20 g/t R-208: 10 g/t M2030; 25 g/t MIBC
3	86.4	6.8	300 g/t Na_2CO_3; 25 g/t R-208; 250 g/t NaHS	110 g/t AX-350; 250 g/t $CuSO_4$
3A	90.0	6.6	125 g/t AX-350 1000 g/t NaHS	
4	85.5	6.7	190 g/t AX-350; 10 g/t M2030; 55 g/t MIBC	50 g/t R-208 250 g/t $CuSO_4$
5A	85.5	6.7	200 g/t AX-350; 10 g/t M2030; 30 g/t MIBC	45 g/t R-208 250 g/t $CuSO_4$
17	85.5	6.7	200 g/t AX-350; 10 g/t M2030; 30 g/t MIBC	45 g/t R-208 250 g/t $CuSO_4$

Table No 2 Flotation Test Results

Test No	Products	Weight %	Assays: g/t, %				% Distribution			
			Au	S	As	Fe	Au	S	As	Fe
1	6 min Rougher Conc	6.74	54.40	18.9	16.18	-	73.9	40.7	59.9	-
	9 min Rougher Conc	9.30	40.40	17.2	15.47	-	75.7	51.0	79.1	-
	Rougher Tailing	90.70	1.33	1.69	0.42	-	24.3	49.0	20.9	-
	Head (calc)	100.00	4.96	3.13	1.82	-	100.0	100.0	100.0	-
2	5 min Rougher Conc	5.81	23.83	9.33	1.96	-	22.1	15.2	6.1	-
	10 min Rougher Conc	14.33	33.50	13.4	9.71	-	76.6	54.0	74.6	-
	13min Rougher Conc	18.40	30.10	13.0	8.87	-	88.2	67.4	87.4	-
	16 min Rougher Conc	22.70	25.20	12.5	7.43	-	91.2	79.5	90.4	-
	Rougher Tailing	77.30	0.72	0.94	0.23	-	8.8	20.5	9.6	-
	Head (calc)	100.00	6.28	3.56	1.86	-	100.0	100.0	100.0	-
3	3 min Rougher Conc	5.65	37.47	16.40	8.38	-	35.9	27.7	25.1	-
	6 min Rougher Conc	12.66	37.95	17.34	11.88	-	81.5	65.5	79.7	-
	9 min Rougher Conc	18.15	29.41	16.21	9.20	-	90.6	87.7	88.5	-
	12 min Rougher Conc	21.29	25.54	14.41	7.98	-	92.3	91.5	90.1	-
	24 min Rougher Conc	24.83	22.43	12.65	6.93	-	94.5	93.7	91.2	-
	Rougher Tailing	75.17	0.43	0.28	0.22	-	5.5	6.3	8.8	-
	Head (cacl)	100.00	5.89	3.35	1.89	-	100.0	100.0	100.0	-
4	Comb Rougher Conc	23.16	26.40	-	-	-	91.2	-	-	-
	Rougher Tailing	76.84	0.77	-	0.28	-	8.8	-	-	-
	Head (calc)	100.00	6.70	-	-	-	100.0	-	-	-
5A	Rougher Conc	24.82	22.50	11.00	6.44	23.90	91.7	87.5	90.6	49.2
	Rougher Tailing	75.18	0.67	0.52	0.22	8.16	8.3	12.5	9.4	50.8
	Head (calc)	100.00	6.08	3.12	1.76	12.10	100.0	100.0	100.0	100.0
17A	Rougher Conc	21.84	24.40	12.90	7.24	24.10	89.0	76.8	89.4	47.0
	Rougher Tailing	78.16	0.84	1.09	0.24	7.58	11.0	23.2	10.6	53.0
	Head (calc)	100.00	5.99	3.67	1.77	11.20	100.0	100.0	100.0	100.0

Figure NO. 1 - Grade-Recovery Relationship



The results indicated that 22-25% of the ore weight was recovered in a sulphide concentrate which assayed 22-24 g/t Au, 11-13% S, 6-7% As & 24% Fe. Gold recovery ranged from 89 to 92% and the flotation tailing assayed 0.7-0.8 g/t Au. Figure No. 1 illustrates grade-recovery relationship (for Tests 2 and 3).

3.2 Treatment of Flotation Concentrate

3.2.1 Cyanidation

The recovery of gold from the flotation concentrate was investigated by cyanidation. The effect of fineness of grind on the recovery of gold was investigated: Test No. 5 was conducted without regrinding prior to cyanidation, Test No. 7 with a "fine" regrind and Test No. 11 with an "ultra-fine" regrind. One test (No. 6) was conducted on the whole ore for comparative purposes.

The pulps were aerated with lime at pH 10-11 for 8 hours before cyanidation in bottle tests on rolls at 33% solids, 2 g/L NaCN, pH 10-11 for 24 to 48 hours.

The test results are summarized in Table No. 3.

Table No. 3 - Cyanidation Test Results

Test No.	Feed	Grind	Reagent Cons, kg/t NaCN	Reagent Cons, kg/t CaO	% Recovery, Au 24 hr	% Recovery, Au 48 hr	Residue g/t Au	Head (calc) g/t Au	Head (direct) g/t Au
5	1st Ro. Conc. from T-4	74% -200 mesh	12.0	13.2	29.2	-	23.3	33.0	33.5
7	Comb. Ro. Conc. from T-4	97% -400	7.04	16.9	39.3	40.8	16.8	28.4	24.4
11	Comb. Ro. Conc. from T-4	92% -11µm	11.1	19.1	39.6	40.8	15.2	25.7	26.4
6	Whole ore	85% -200 mesh	9.50	9.06	-	33.0	3.83	5.71	5.71

Cyanidation tests on the whole ore and concentrate indicated that the sample was refractory. Very fine grinding prior to cyanidation did not improve the results. Gold recovery from the flotation concentrate was 41% and from the whole ore only 33%.

3.2.2 Pressure Oxidation

Pressure oxidation tests were conducted to examine the oxidation of sulphides and subsequent recovery of gold by cyanidation and CIL.

Figure No. 2 illustrates the basic flowsheet used in a test program, which involved regrinding of flotation concentrate followed by pressure oxidation, filtration of the autoclave discharge pulp, neutralization of the autoclave solids and cyanidation (CIL) of the neutralized solids.

a. Pressure Oxidation Tests

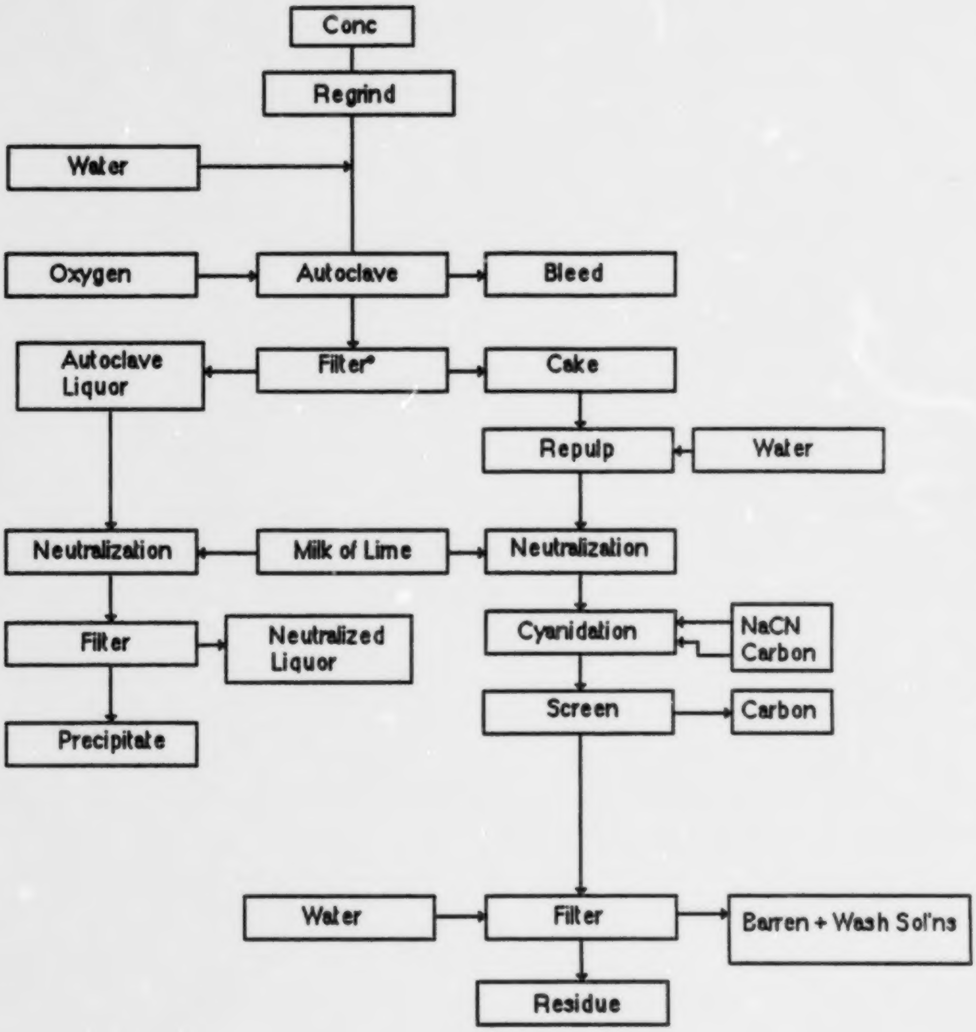
The pressure oxidation tests were conducted in a 2 L titanium Parr autoclave.

The test conditions were as follows:

Temperature: 200°C
Time: 90-120 min
Oxygen Pressure: 100 psi
L/S Ratio: 5/1
Feed: flotation concentrate
Fineness of Grind: 97% minus 400 mesh (cy residue)

At the end of each test the autoclave discharge pulp was filtered, the liquor was submitted for analysis and neutralization and the residue for neutralization and cyanided. The filtration characteristics of the autoclave pulps were good.

Figure No. 2 - Flowsheet


Conc
Regrind
Water
Oxygen
Autoclave
Bleed
Autoclave Liquor
Filter*
Cake
Repulp
Water
Neutralization
Milk of Lime
Neutralization
Filter
Neutralized Liquor
Cyanidation
NaCN
Carbon
Screen
Carbon
Precipitate
Water
Filter
Barren + Wash Sol'ns
Residue

*No wash

The autoclave test conditions and results are presented in Table No. 4.

Table No. 4 - Autoclave Test Conditions and Results

Test No.	Feed	Oxidation Time, min.	Autoclave Solution Analysis				
			Emf, mV	Free Acid (g/L, H_2SO_4	Fe^{3+}, mg/L	Fe^{2+}, mg/L	As mg/L
8	Ro. Conc. (Test 4)	90	520	45.0	5010	<10	940
9	Comb. Ro. Conc. (T-4)	90	600	29.0	2375	125	NA
10	Comb. Ro. Conc. (T-4)	120	610	32.0	2416	34	267

The following observations were made:

1. Sulphide oxidation degree is the most important measurement in the evaluation of pressure oxidation data. Autoclave residues in all three tests assayed less than 0.2% $S^{=}$, which was more than 98% sulphide oxidation. This suggested that most of the available gold was exposed for recovery by cyanidation.

2. The production of elemental sulphur was negligible.

3. In regard to environmental concerns, arsenic was precipitated during neutralization as ferric arsenic, $FeAsO_4$. Considerable attention was given to the subsequent stability of this compound which was evaluated as described in "3.2.2" section of this report.

The arsenic concentration in the autoclave liquor ranged from 900 to 300 mg/L. The iron to arsenic ratio was in excess of 5:1 in solution ahead of precipitation.

b. Cyanidation (CIL) Tests

The autoclave filter cake was repulped with water to 30% solids and neutralized with lime slurry to pH 11. The tests were conducted in bottles on rolls. NaCN concentration was maintained at 2 g/L. Carbon concentration was 10 g/L. Total retention time was 24 hours. To examine leaching rates, loaded carbon was replaced with fresh carbon after 6 and 16 hours of agitation.

Cyanidation test results are presented in Table No. 4.

Table No. 4 - Cyanidation Results

Test No.	Reagent Cons, kg/t*		% Recovery, Au			Residue					Head (calc) g/t Au
	NaCN	CaO	6h	16 h	24h	Weight %*	Assays				
							Au, g/t	$S^=$, %	As, %	Fe, %	
8A	2.40	48.0	NA	NA	95.9	96.0	1.07	-	-	-	24.8
9A	6.73	39.7	89.8	93.8	93.9	93.5	1.58	0.25	-	-	24.0
10A	7.48	36.8	96.1	97.2	97.4	93.2	0.63	0.56	7.9	20.5	22.5

* Autoclave feed

The results indicate that:

1. The highest gold extraction of 97% was achieved, when the pressure oxidation test was extended to 120 minutes.

   The final residue assayed 0.6 g/t Au and 0.06% $S^=$.

2. The leaching process was completed after 6-16 hours of agitation.

3. NaCN consumption averaged 5.5 kg/t autoclave feed or 1.4 kg/t ore.

4. The average cyanide residue represented 94 weight % of the autoclave feed.

3.2.3 Pressure Oxidation with Nitric Acid

Four nitrate catalyzed oxygen pressure leach tests were conducted for comparison with the "conventional" pressure oxidation tests. The same flowsheet shown in Figure No. 2 was used. Due to the limited test program, only preliminary amenability tests were conducted.

The test conditions are shown below:

Temperature:	80°C
Time:	60-90 min.
Oxygen Pressure:	100 psi
L/S Ratio:	10/1-17/1
HNO_3 concentration:	20.8-62.5 g/L
Feed:	flotation concentrate
Fineness of Grind:	97% minus 400 mesh (cy residue)

During the testwork campaign some difficulties were experienced in maintaining the oxygen pressure and controlling the heat generated by the exothermic leaching reaction. Overheating above 120°C (the melting point of S°) caused formation of sulphur-sulphide pellets which contained gold and unreacted sulphides. The conditions and results for the nitrate catalysed oxidation tests and subsequent cyanidation of the autoclave discharge (filtered and neutralized) are summarized in Tables No. 5 and 6.

Table No. 5 - Test Conditions and Results

Test No.	Oxid. Time min.	HNO_3 g/L	% Solids	O_2 Pres. psi	Autoclave Solution Analysis					Comments
					Emf mV	Free Acid g/L H_2SO_4	Fe^{3+} mg/L	Fe^{2+} mg/L	As mg/L	
12	60	62.5	9.1	100	>700	4.7	-	-	-	$S^\circ/S^=$ pellets
14	60	62.5	9.1	300	>700	27.7	18233	167	6480	$S^\circ/S^=$ pellets
16	90	20.8	9.1	100	>700	6.6	10780	>10	2790	-
17	60	20.8	6.0	100	>700	11.0	7625	<10	2530	-

Table No. 6 - Cyanidation Test Results

Test No.	Reagent Cons., kg/t* NaCN	CaO	% Recovery Au	Weight %*	Residue Au g/t	Head g/t Au
16	7.44	29.0	83.2	63.4	5.10	19.2
17	7.07	29.3	89.7	70.0	3.80	25.8

* autoclave feed

** autoclave + cyanidation

The following observations were made:

1. Additional testwork is required to develop optimum condtions. HNO3 concentration should be reduced to 5-10 g/L to control the rate of reaction and elemental sulphur formation.

2. The autoclave liquors assayed about 0.5 mg/L Au which accounted from 26 to 30% of the gold in the autoclave feed. Previous experience has indicated that gold recovery from acid liquor is very efficient with carbon adsorption.

3.2.4 Roasting

The flotation concentrate was roasted in two stages in a muffle furnace. The roasting conditions were as follows:

Feed:	flotation concentrate
Total Roasting Time:	90-180 min
1st Stage Roast Time:	30 min
2nd Stage Roast Time:	60 - 150 min
Maximum Temperature:	600-665°C
Soaking Temperature:	580-590°C

The roasted calcine was cyanided in bottles on rolls with 2 g/L NaCN for 24 hours. The test conditions and results are contained in Table No. 7.

The gold recovery achieved in Test 15 was 79%, leaving a residue assaying 5.3 g/t Au; 2.1% As and 0.03% $S^{=}$. To decompose the arsenates in the leach residue from Test 15 and to determine the quantity of gold occluded by the arsenates, the residue was filtered, repulped and leached in 150 g/L NaOH solution at 20% solids for 2 hours at 100°C. The pulp was filtered, washed, repulped and leached with 10 g/L NaCN for 24 hours.

The results indicated that 33% of the gold was associated with arsenates. The final residue contained 3, 8 g/t Au and 0.5% As. Overall gold extraction in Test 15 (including the NaOH leach) was 86%.

Table No. 7 - Roasting Conditions and Results

Test	Conc.	Roasting				Cyanidation						Head
No.	Assay Au, g/t	Time Min.	Temp. °C	Calcine, wt.		Reagent Cons., kg/t		Residue Assays			% Au Ext'n. 24 hrs.	(calc) Au g/t
				grams	%	NaCN	CaO	Au, g/t	$S^{=}$, %	As, %		
13	22.5	180	580	225	82.6	2.82	8.47	7.24	0.023	NA	71.2	25.2
15	22.5	90	590-665	225	84.0	2.42	11.6	5.31	0.032	2.10	79.2	25.5

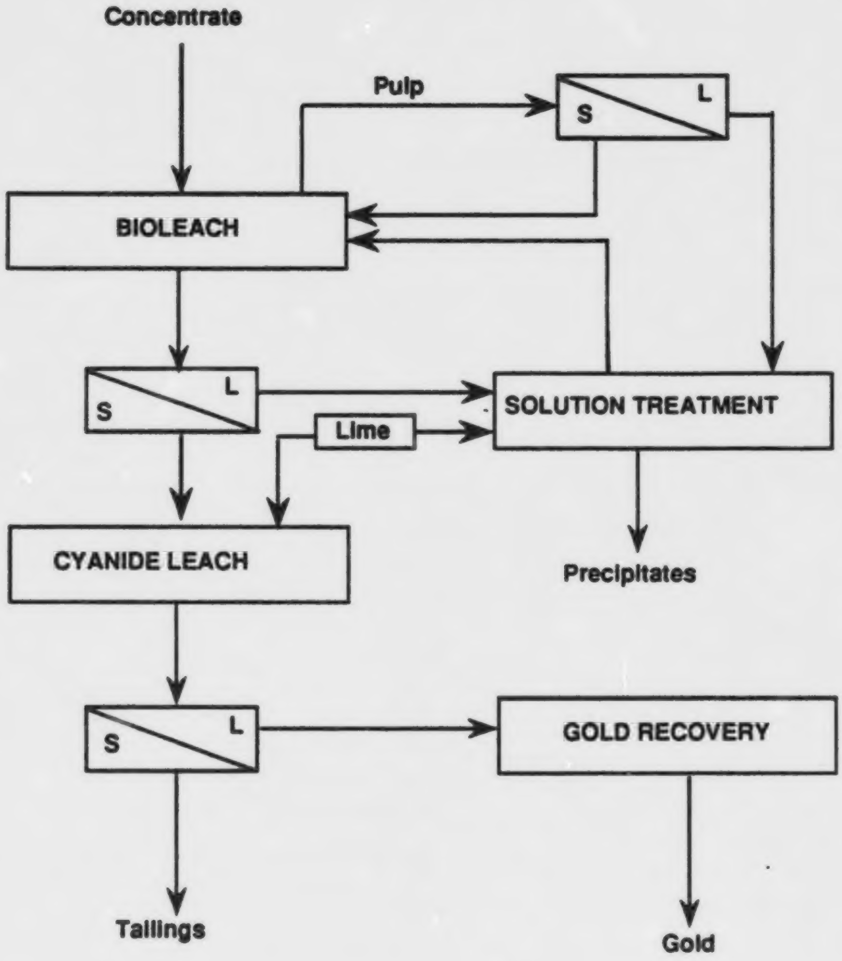
3.2.5 Biological Oxidation

Biological leaching was evaluated as one of the methods for gold extraction from the flotation concentrate. The main objectives of this batch test were:

- To determine if the concentrate is bioleachable.
- To evaluate gold recovery data as a function of iron oxidation.
- To generate data in order to establish the extraction-retention time relationship.
- To estimate the rate of oxidation.

Figure No. 3 shows a suggested schematic flowsheet of a gold recovery circuit for refractory concentrates incorporating biological oxidation.

Figure No. 3 - Biological Oxidation for Refractory Gold Concentrate.


Concentrate
Pulp
S
L
BIOLEACH
S
L
SOLUTION TREATMENT
Lime
CYANIDE LEACH
Precipitates
S
L
GOLD RECOVERY
Tailings
Gold

3.2.5.a. Laboratory Batch Test Procedure

"Thiobacillus Ferro-oxidans" bacteria were used in the testwork. Flotation rougher concentrate from Test 17 ground to 97% minus 400 mesh was tested.

The following independent variables were monitored and adjusted during the test program:

1. Temperature: 35-38°C
2. Dissolved Oxygen Concentration: > 2 mg/L
   An air sparge enriched with CO2 was applied
3. pH: 1.2-2.0
4. EMF
5. Liquor Specific Gravity
6. Solids concentration

To acclimatize the bacteria (developed on similar feed) to the arsenopyrite concentrate, the culture was added to 100 grams of reground concentrate and mixed with water at 10% solids. The mixture was sparged with an air/CO_2 mixture and the following reagents were added.

20 mL H_2SO_4 (98%) to pH = 1.2
5 g/L $Fe_2(SO_4)_3$ - (commercial grade)
2 g/L $FeSO_4$ - (reagent grade)
9K media: 2 g/L $(NH4)_2SO_4$
0.5 g/L Mg SO_4
2.0 g/L $KHPO_4$

The above conditions were maintained for 5 days after which 300 mL of slurry were inoculated into the reactor. The reactor was a 15L baffled container. Agitation was provided by a mechanical mixer with a turbine propeller. An air-CO_2 mixture was sparged and evenly distributed through the slurry by a specially designed oxygen diffuser. The reactor was placed in a water bath equipped with an immersion heater.

1 kg of ground concentrate was mixed with water to 10L pulp volume. The volume was held constant to correct for evaporation and sampling losses. Baseline data, which included pH, temperature, EMF, liquor specific gravity, was obtained. Liquor samples were removed and analyzed for Fe^{2+}, Fe^{3+} and As. Daily operating data is presented in Table No. 8.

Table No. 8 - Bioleach Daily operating Data .

Date	Oxygen Dissolved (mg/L)	Oxygen Uptake (mg/L/hr)	pH	Temp. (C)	EMF (mV)	Liquor S.G. (g/mL)	ASSAYS, mg/L Fe 2+	Fe 3+	FeTotal	As	Fe(AsPy)	Fe (Py)	Py Oxidation Rate g/L/day	Comments
88/12/09	-	-	1.96	30	425	-	4280	302	4782	362	270	4512		100 ml Thiobacillus sp. + nutrients added
88/12/12	-	-	2.11	30	470	-	4784	1816	6600	441	329	6271		
88/12/13	-	-	2.23	32	460	1.014	4621	779	5400	222	166	5234		
88/12/14	-	-	2.12	34	435	1.012	5388	312	5700	257	192	5508		
88/12/15	-	-	2.12	34	490	1.015	7158	342	7500	292	218	7282	1.77	
•			-	34	-	-	4956	2644	7600	-	-	-	-	after increasing mixer speed and aeration
88/12/16	5.90	90	1.15	34	505	1.024	3251	4847	8098	908	677	7421	0.96	remove 1L pulp for Cyn-1
88/12/17	5.95	55	1.20	34	505	1.026	3646	4951	8597	-	-	-	-	
88/12/18	4.92	39	1.35	35	505	1.035	4672	6288	10960	-	-	-	-	
88/12/19	4.92	34	1.35	35	505	1.040	4225	6115	10341	-	-	-	-	300 ml innoculation from beaker + 100 ml from Dickenson stored culture
•														
88/12/20	8.04	19	1.39	28	505	1.042	4649	3699	12140	1950	1454	10686	0.86	
88/12/21	5.71	50	1.25	34	505	1.056	3646	12306	15952	3340	2490	13462	1.14	
88/12/22	5.87	68	1.34	35	565	1.052	1025	12192	13217	-	-	-	-	remove 1L pulp for Cyn-2
88/12/23	5.79	72	1.29	35	645	1.046	285	14470	14755	-	-	-	-	remove 1L pulp for Cyn-3
88/12/24	5.84	80	1.34	34	700	1.054	<10	13600	13610	6030	4496	9114	0.36	remove 1.4 L pulp for Cyn-4
88/12/25	6.80	66	1.28	33	<700	1.046	<10	13070	13080	4250	3169	9911	0.40	remove 2L pulp for Cyn-5
88/12/26	7.59	68	1.28	35	<700	1.052	<10	13260	13270	4330	3229	10041	0.38	remove 3L pulp for Cyn-6
88/12/28	6.50	64	1.30	35	<700	1.052	<10	13400	13410	4380	3266	9030	0.25	

3.2.5.b Cyanidation

To evaluate the recovery of gold at various degrees of iron oxidation, pulp samples were periodically removed, filtered and washed. The solids were repulped with water, neutralized with lime to pH = 11.0 and cyanided in bottles on rolls for 24 hours with 2 g/L NaCN. Test results are summarized in Table No. 9.

Table No. 9 - Cyanidation Test Results

Time days	Test No	Reagent Cons, kg/t		Residue Assays				% Rec'y Au	Calc Head Au g/t
		NaCN	CaO	Au, g/t	S=, %	Fe, %	As, %		
0	6	11.06	19.06	15.20	-	24.1	7.24	40.9	26.4
8	cyn-1	9.20	20.61	11.13	-	-	-	54.3	24.4
14	cyn-2	6.49	25.20	7.68	2.29	11.2	2.96	67.8	23.4
15	cyn-3	6.78	16.90	4.68	1.48	10.8	1.80	84.0	29.3
16	cyn-4	6.84	12.40	4.17	0.96	9.67	1.14	85.7	29.0
17	cyn-5	6.37	15.80	3.75	0.38	9.02	0.88	87.8	30.8
18	cyn-6	7.53	17.80	2.72	0.30	8.63	0.71	90.4	28.5

3.2.5.c. Neutralization of Bioleach Liquor

In preparation for disposal, 500 mL of final liquor was neutralized with lime slurry in two stages:

to pH = 5 with 24.8 kg/t $Ca(OH)_2$ for 1 hour

to pH = 11 with 15.6 kg/t $Ca(OH)_2$ for 1 hour.

The slurry was filtered and the precipitate was dried. Wet and dry weights were recorded. Neutralization of 500 mL of liquor produced 58 g (dry) precipitate. The wet precipitate contained 35% moisture.

No further testwork was performed on the "waste" products of biological leaching.

3.2.5.d. Discussion of Batch Test Results

Iron extraction was calculated using the following formula:

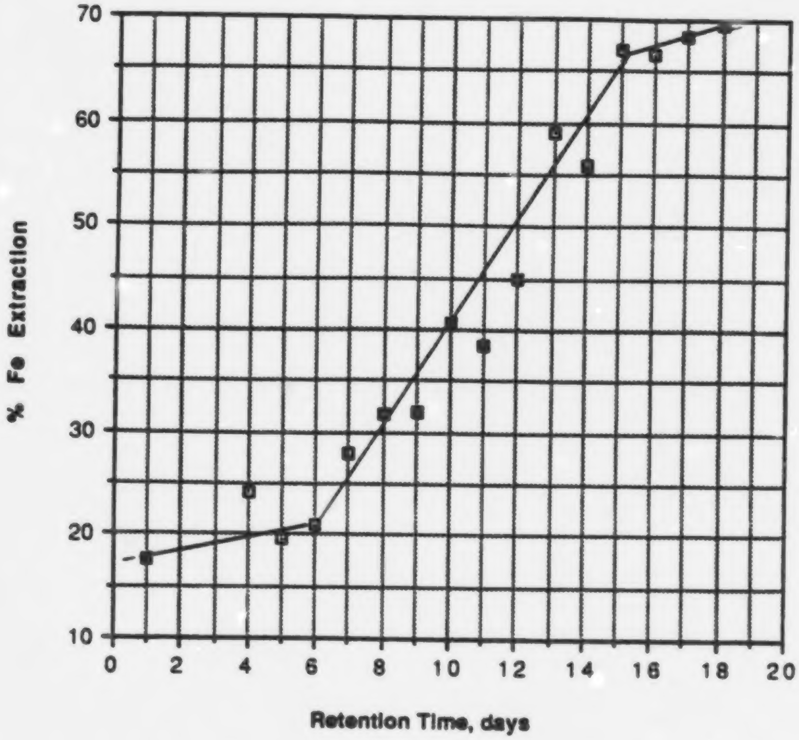
$$\% \text{ Fe Extraction} = \frac{\text{Fe in solution (mg/L) X Solution Volume (L)}}{\text{Fe in Solids (\%) X Sample Wgt (mg)}} \text{ X } 100$$

The following assumptions for the calculations were made:

- all sulphides were present as arsenopyrite and pyrite
- no weight reduction occurred
- no change in the solids specific gravity occurred. Solids S.G. was 3.32.

The iron extraction curve is shown in Figure No. 4. It is a typical extraction curve with a lag phase, biological growth period and dormant phase.

Figure No. 4 - Retention time-Iron Extraction Relationship.

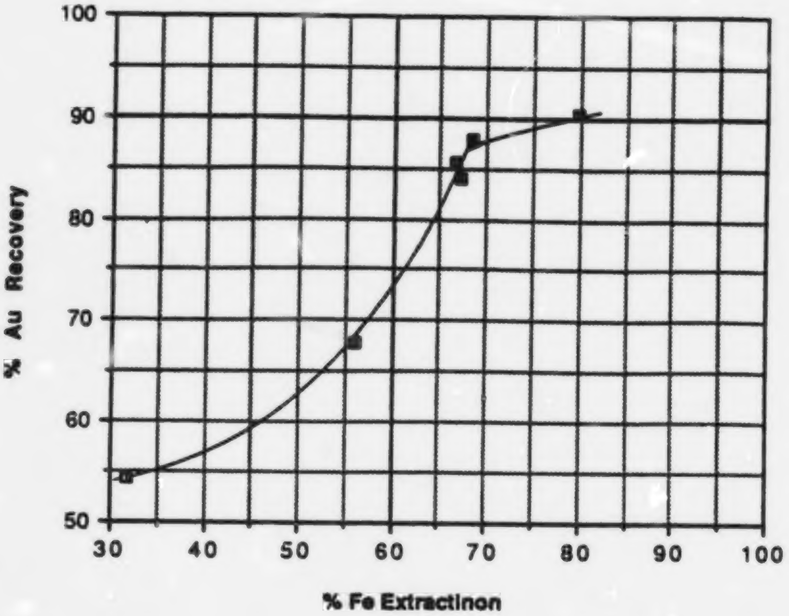


The relationship between iron extraction and gold recovery is illustrated in Figure No. 5 and in Table No. 10.

Table No. 10 - Fe Extraction - Gold Recovery Relationship

Time days	% Fe Extraction	% Au Recovery
0	-	40.9
1	17.5	-
4	24.2	-
5	19.6	-
6	21.0	-
7	27.8	-
8	31.8	54.3
9	31.9	-
10	40.6	-
11	38.4	-
12	45.0	-
13	59.1	-
14	56.0	67.8
15	67.1	84.0
16	66.7	85.7
17	68.5	87.8
18	69.6	90.4

Figure No. 5 - Fe Extraction vs Gold Recovery



The results indicated that 90% gold extraction was achieved with 70% iron oxidation.

The degree of iron sulphide oxidation was measured by iron extraction. The rate of pyrite oxidation ranged from 1.77 to 0.25 g FeS_2/L/day and is shown in Table No. 8.

The Fe^{3+}/Fe^{2+} ratio achieved was >50/1. The total duration of the test was considered as the residence time for the batch test. Continuous tests would have to be conducted to determine operating retention times. Due to the sampling procedure etc., the final residue (biomass) weight could not be precisely determined. It was estimated that the final residue represented about 75-85 wgt. % of the test feed. The final residue assayed 32 g/t Au, 0.8% $S^=$, 4.4% SO_4.

From the test results it was concluded that the flotation concentrate could be bioleached. Continuous testing is recommended for equipment sizing and cost analysis.

3.3 Environmental Studies

In preparation for the disposal of waste streams, the stability of the arsenic compounds, which were precipitated during neutralization, was examined. Only pressure oxidation products i.e., autoclave liquor and final CIL residue were tested.

3.3.1 Chemistry

An overview of the chemistry of the arsenic compounds in the autoclave discharge and neutralized autoclave solution is outlined below.

3.3.1.a Products in Autoclave Discharge

Following destruction of arsenopyrite and any other arsenic minerals in high temperature (+180°C) oxidative process, the slurry is reported to contain the following range of arsenical entities:

liquor: H_3AsO_4, $H_2AsO_4^-$, $HAsO_4^{2-}$, $AsO_4 3^-$
solids: $FeAsO_4.xH_2O$ (Robins -6)
: $Fe_2(AsO_4)\ SO_4)\ OH.nH_2O$ (Weir & Berezowski -3)
: variants of these compounds
: Fe^{2+} compounds are not well documented

The high temperature forms of ferric arsenate are reported to be relatively stable in acidic solution (Krause and Ettel -2) as indicated by the soluble arsenic concentration in water after extended exposure.

Synthetic scorodite $FeAsO_4 . 2H_2O$
- >0.5 mg/L As in pH range 2.4 - 5.0
Natural scorodite $FeAsO_4 . 2H_2O$
- ~0.4 - 3.6 mg/L As at pH 5.0
but the arsenic increases significantly at alkaline pH.

( ) References listed in 3.3.1.c.

3.3.1.b. Precipitation of Arsenic from Autoclave Solution

"Basic ferric arsenate" precipitates - $FeAsO_4 . x Fe(OH)_3$ - prepared at pH 5 were found by Krause and Ettel to be increasingly stable as the Fe/As molar ratio increased from 4 to 17.

Fe/As Ratio	Stability Range pH *
4	3 - - 5.3
8	3 - - 8.2
17	3 - - 9.0

* <0.5 mg/L in solution on long contact with water at indicated pH.

Robins (6) has indicated that thermodynamic data show a limited range of stability for $FeAsO_4$ at pH 2-3, and that above this pH it breaks down to goethite. Krause and Ettel (1) confirmed this interpretation, but conclude that the "basic ferric arsenate" with high Fe/As ratios is relatively stable (see above data).

Arsenic precipitated at pH 5 in the presence of high ratios of ferric iron does not appear to form a calcium compound but calcium arsenate which can form over a wide pH range is relatively unstable at high pH. In the presence of small quantities of CO_2 it dissociates to $CaCO_3$, remobilizing the arsenic (Robins and Tozawa - 5). Consequently the formation of $Ca_3(AsO_4)_2$ is to be avoided, but these authors point out that the efficiency of lime additions in removing arsenic from alkaline gold solutions may be due to the formation of more stable arsenates with the residual base metal ions.

3.3.1.c. References

1) E. Krause, V. Ettel - Ferric Arsenate Compounds; Solubilities of Basic Ferric Arsenates - Impurity Control and Disposal, Proc. 15th Annual Hydromet. Meeting C.I.M. Vancouver, August 1985.

2) E. Krause, V. Ettel - Solubilities and Stabilities of Ferric Arsenates, Proc. 17th Annual Hydromet. Meeting, C.I.M., Saskatoon 1987

3) D.R. Weir, R. Berezowsky - Aqueous Pressure Oxidation of Refractory Gold Feed Stocks - 26th Conference of Metallurgists, C.I.M., Winnipeg, August 1987

4) J. Dutrizac, J. Jambor - The Synthesis of Crystalline Scorodite, $FeAsO_4$, 2 H_2O - Hydrometallurgy, 19 (1987) 377-384, Elsevier, Amsterdam

5) R. Robins, K. Tozawa - Arsenic Removal from Gold Processing Waste Waters: The Potential Ineffectiveness of Lime, C.I.M. Bulletin, April 1982, p. 171-174

6) R. Robins - The Aqueous Chemistry of Arsenic in Relation to Hydrometallurgical Processes: Impurity Control and Disposal, Proc. 15th Annual Hydromet. Meeting C.I.M., Vancouver, August 1985.

3.3.2 Stability of Testing

Flowsheets for the arsenic stability testing on neutralized autoclave liquor and CIL residue are shown in Figures No. 6 and 7. The autoclave liquor was neutralized with lime slurry in two stages to pH = 5 for 1 hour and to pH = 11.0 for 1 hour.

The neutralized slurries were filtered, washed three times and repulped with distilled water to 50 g/L dry solids. The slurries were placed in glass bottles and the pH was adjusted and maintained at 8 with H_2SO_4. The samples were repeatedly agitated for the duration of 180 hours.

At the end of the tests, the slurries were filtered and the products were submitted for analysis. (The solution was further filtered through a 0.45 μm nitrocellulose membrane prior to analysis.)

The test conditions and results are presented in Table No. 11.

The results indicated that at pH 8 the equilibrium arsenic concentration was less than 2 mg/L.

Table No. 11 - Arsenic Compound Stability Date

1. Conditions

Feed	Test	Neutralization		Test	H2SO4 Req't
		pH	CaO, g/L*	pH	kg/t**
CIL Residue T-10	10-A1	11	-	8	10.7
	10-A2	11	-	8	5.6
Autoclave Liquor T-10	10-B1	5	22.7		
		11	2.0	8	34.7
	10-B2	5	17.7		
		11	7.6	8	35.8

* Autoclave Liquor
** Neutralized Residue

2. Results

Test No	Feed Assay		180 h Solution Analysis, mg/L		Solids g/L	Residue Assays %	
	Fe	As	Fe	As		Fe	As
10-A1	20.5%	7.89%	1.55	7.90	45	23.1	8.51
10-A2	20.5%	7.89%	1.62	9.97	45	23.3	8.51
10-B1	2450mg/L	267mg/L	0.05	0.41	30	2.38	0.27
10-B2	2450mg/L	267mg/L	0.17	0.14	24	2.71	0.31

Figure No. 6

Flowsheet

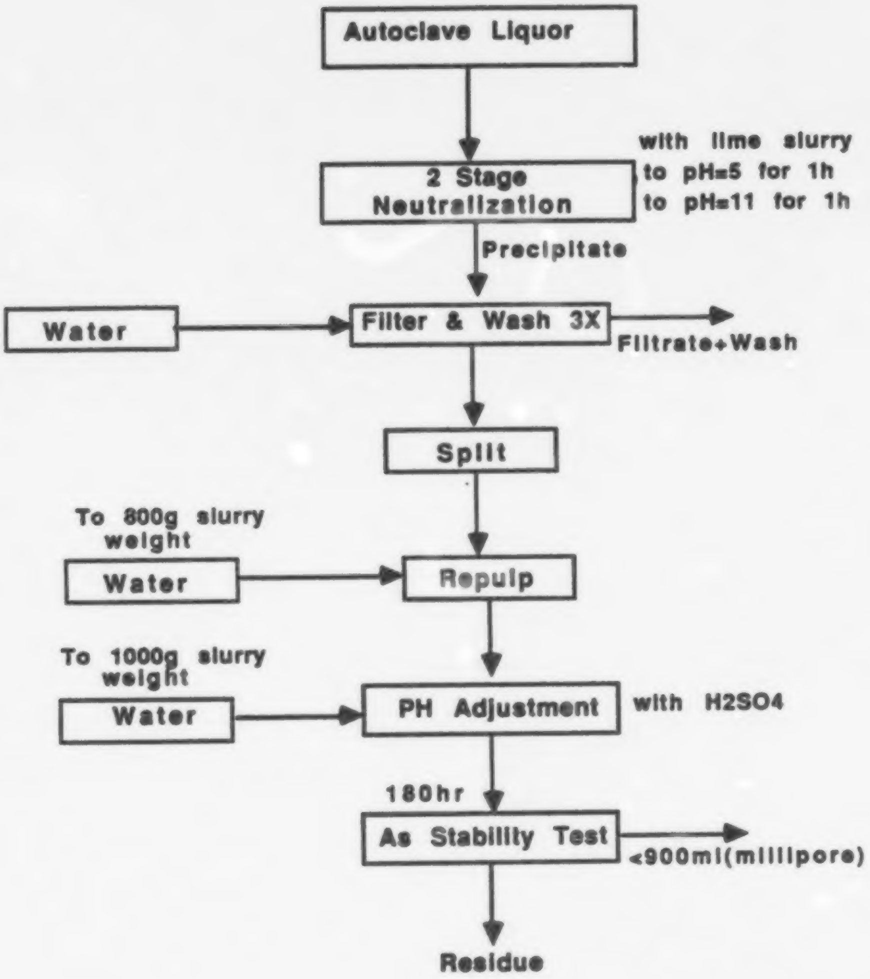
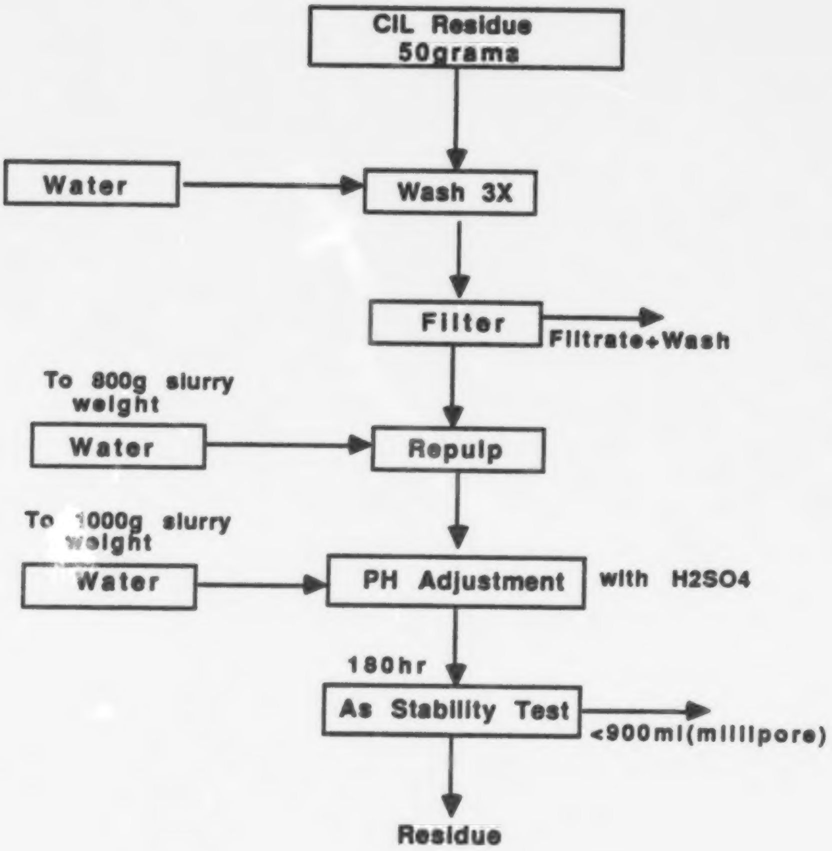

Autoclave Liquor
2 Stage Neutralization
with lime slurry
to pH=5 for 1h
to pH=11 for 1h
Precipitate
Water
Filter & Wash 3X
Filtrate+Wash
Split
To 800g slurry weight
Water
Repulp
To 1000g slurry weight
Water
PH Adjustment
with H2SO4
180hr
As Stability Test
<900ml(millipore)
Residue

Figure No. 7

Flowsheet


CIL Residue 50grams
Water
Wash 3X
Filter
Filtrate+Wash
To 800g slurry weight
Water
Repulp
To 1000g slurry weight
Water
PH Adjustment
with H2SO4
180hr
As Stability Test
<900ml(millipore)
Residue

SAMPLE PREPARATION

On October 24, 1988, 600 kg of ore was received at Lakefield Research.

The ore was crushed to minus 1/2 inch. 1/4 of ore was riffled out and further crushed to minus 10 mesh. Head sample was removed and test charges were prepared. The remaining ore was stored.

DETAILS OF TESTS

Test No. 1

Purpose:	To conduct a sulphide rougher test on a Aspy - Po
Procedure:	As indicated below
Feed:	2000 grams of -10 mesh (Aspy - Po ore)
Grind:	2 kg ground in the lab ball mill at 65% solids for 30 minutes.
Conditions:	

Stage	Reagents Added, grams per tonne			Time, minutes			pH
	A350	R208	MIBC	Grind	Cond.	Froth	
Rougher 1	20	20	20	-	3	3	6.7
Rougher 2	15	15	-	1	3	-	
Rougher 3	10	10		-	1	3	

Stage	Rougher
Flotation Cell	1000 g D-1
Speed: r.p.m.	1800

Observations: Slow Sulphide floatation
2nd Ro. Conc. was examined under binocular microscope.
- fine grained sulphides (black)
- Aspy, Py
- Reddish goethite

3608

Test No. 1

Metallurgical Results

Product	Weight %	Assays: %, g/t			% Distribution		
		Au	S	As	Au	S	As
1. Rougher Conc 1	6.74	54.4	18.9	16.18	73.9	40.7	59.9
2. Rougher Conc 2	2.56	3.41	12.6	13.6	1.8	10.3	19.1
3. Rouhger Tail	90.70	1.33	1.69	0.42	24.3	49.0	20.9
Head (calc)	100.0	4.96	3.13	1.82	100.0	100.0	100.0

Calculated Grades and Recoveries

1+2	9.30	40.36	17.17	15.47	75.7	51.0	79.1

Screen Analysis

Mesh Size (Tyler)	% Retained Individual	% Retained Cumulative	% Passing Cumulative
+ 65	0.9	0.9	99.1
100	4.6	5.5	94.5
150	12.8	18.3	81.7
200	16.9	35.2	64.8
270	13.2	48.4	51.6
400	8.9	57.3	42.7
- 400	42.7	100.0	-
Total	100.0	-	-

Test No. 2

Purpose: To repeat test No. 1, but with a finer grind, longer flotation times and add $CuSO_4$ to rougher No. 3.

Procedure: Grind and float a series of rougher concentrates.

Feed: 2000 grams of minus 10 mesh ASPY - Po ore.

Grind: 45 minutes at 65% solids in the lab ball mill.

Conditions:

	Reagents Added, grams per tonne					Time, minutes			pH
	A350	R208	MIBC	$CuSO_4$	M2030	Grind	Cond.	Froth	
Grind	-	-	-	-	-	45	-	-	-
Ro. Conc. No.1	25	10	15	-	-	-	2	5	6.5
Ro. Conc. No. 2	15	5	5	-	-	-	1	5	-
Condition	-	-	-	250	-	-	3	-	-
Ro. Conc. No. 3	20	5	5	-	-	-	1	3	5.5
Ro. Conc. No. 4	20	-	-	-	10	-	1	3	-

Metallurgical Results

Product	Weight %	Assays %			% Distribution		
		Au	S	As	Au	S	As
1. Ro. Conc. No. 1	5.81	23.83	9.33	1.96	22.1	15.2	6.1
2. Ro. Conc. No. 2	8.52	40.17	16.2	15.0	54.5	38.8	68.5
3. Ro. Conc. No. 3	4.07	17.84	11.7	5.88	11.6	13.4	12.8
4. Ro. Conc. No. 4	4.30	4.34	10.0	1.30	3.0	12.1	3.0
5. Ro. Flot. Tail	77.30	0.72	0.94	0.23	8.8	20.5	9.6
Head (Calc)	100.00	6.28	3.56	1.86	100.0	100.0	100.0

Calculated Grades and Recoveries

Ro. Concs. 1+2	14.33	33.5	13.4	9.71	76.6	54.0	74.6
Ro. Concs. 1-3	18.40	30.1	13.0	8.87	88.2	67.4	87.4
Ro. Concs 1-4	22.70	25.2	12.5	7.43	91.2	79.5	90.4

Screen Analysis - Rougher Tail #2A

Mesh Size (Tyler)	% Retained		% Passing Cumulative
	Individual	Cumulative	
+ 65	0.1	0.1	99.9
100	0.4	0.5	99.5
150	3.1	3.6	96.4
200	10.0	13.6	86.4
270	14.8	28.4	71.6
400	14.4	42.8	57.2
- 400	57.2	100.0	-
Total	100.0	-	-

MINERALOGY

Rougher Float Tails Test 2

A heavy liquid separation was carried out as a portion of the sample at S.G. 2.9. The sink fraction was mounted in PS 2082 and examined with the ore microscope. The opaque mineral assemblage, listed in decreasing abundance, are:

Marcasite - the product of pyrrhotite alteration, occurs with intermixed iron oxides ± gangue. It is very porous and represents the complete oxidation of pyrrhotite to secondary products.

Goethite and Ilmenite (+Leucoxene) - are both abundant as liberated grains and as middlings with marcasite.

Arsenopyrite and Pyrite - are both minor in occurrence. The former generally forms middling grains with marcasite or iron oxides/hydroxides.

Pyrrhotite and Chalcopyrite - are present in trace amounts only. The few fresh grains of pyrrhotite are generally enclosed in gangue.

Test No. 3

Purpose: To repeat test No. 2 but at a higher pH (add Na_2CO_3 to the grind), add $CuSO_4$ to the 1st stage and increased collector.

Procedure: As indicated below.

Feed: 2000 grams of minus 10 mesh ore.

Grind: 45 minutes at 65% solids in the lab ball mill.

Conditions:

Stage	Reagents Added, grams per tonne						Time, minutes			pH
	Na_2CO_3	$CuSO_4$	AX 350	AP 238	MIBC	NaHS	Grind	Cond.	Froth	
Grind	1500	-	-	-	-	-	45	-	-	6.8
Condition (1)	500	250	-	-	-	-	-	3	-	-
(2)	1000	-	50	10	-	-	-	2	-	6.9
Ro. Conc. No. 1	-	-	-	-	10	-	-	1	3	-
Ro. Conc. No. 2	-	-	15	5	5	-	-	1	3	6.4
Condition	-	-	-	-	-	250	-	3	-	7.0
Ro. Conc. No. 3	-	-	20	5	2.5	-	-	1	3	-
Ro. Conc. No. 4	-	-	25	5	2.5	-	-	1	3	-

Test No. 3

Metallurgical Results

Product	Weight %	Assays: %, g/t			% Distribution		
		Au	S	As	Au	S	As
1. Rougher Conc 1	5.39	37.47	16.4	8.38	35.3	26.0	25.2
2. Rougher Conc 2	6.67	38.33	18.1	14.7	44.7	35.5	54.7
3. Rougher Conc 3	5.23	9.73	13.6	3.03	8.9	20.9	8.8
4. Rougher Conc 4	3.00	3.18	4.02	0.94	1.7	3.5	1.6
5. Rougher Tail	79.72	0.68	0.60	0.22	9.5	14.1	9.8
Head (calc)	100.0	5.72	3.40	1.79	100.0	100.0	100.0

Calculated Grades and Recoveries

1+2	12.06	37.95	17.34	11.88	80.0	61.5	79.8
1 to 3	17.29	29.41	16.21	9.20	88.9	82.4	88.7
1 to 4	20.28	25.54	14.41	7.98	90.5	85.9	90.2

Test No. 3A

Purpose: To investigate the recovery of gold from rougher flotation tailing of Test 3 by flotation at finer grind and with NaHS additions.

Procedure: As indicated below

Feed: 1485 grams of flotation tail from test No. 3

Grind: 1485 grams in the lab ball mill for 10 min. at 50% solids.

Conditions:

Stage	Reagents Added, grams per tonne			Time, minutes			pH
	AX350	NaHS	MIBC	Grind	Cond.	Froth	
Rougher 1	25	-	8	-	1	3	6.4
Rougher 2	-	500	-	-	3	-	-
Rougher 3	-	500	-	-	3	-	-
	50	-	-	-	1	5	6.6

Stage	Rougher
Flotation Cell	1000 g D-1
Speed: r.p.m.	1800

Test No. 3A

Metallurgical Results

Product	Weight %	Assays: %, g/t			% Distribution		
		Au	S	As	Au	S	As
1. Rougher Conc 1	1.16	5.0	2.21	0.61	10.0	7.1	3.2
2. Rougher Conc 2	1.62	3.88	2.49	0.7	10.8	11.2	5.0
3. Rougher Conc 3	1.72	2.84	1.63	0.59	6.4	7.7	4.6
4. Rougher Tail	95.50	0.43	0.29	0.20	70.7	74.0	87.2
Head (calc)	100.0	0.58	0.36	0.22	100.0	100.0	100.0

Calculated Grades and Recoveries

1+2	2.78	4.35	2.37	0.64	20.9	18.3	8.2
1 to 3	4.50	3.77	2.09	0.62	29.3	26.0	12.8

Screen Analysis

Mesh Size (Tyler)	% Retained		% Passing
	Individual	Cumulative	Cumulative
+ 100	0.2	0.2	99.8
150	2.0	2.2	97.8
200	7.9	10.1	89.9
270	13.4	23.5	76.5
400	15.3	38.8	61.2
- 400	61.2	100.0	-
Total	100.0	-	-

Test No. 3 - G'11

Metallurgical Results

Product	Weight	Assays: %, g/t			% Distribution		
	%	Cu	S	Ag	Cu	S	Ag
1 Rougher Conc	5.66	37.47	16.4	8.83	35.9	27.7	25.1
2 Rougher Conc 2	7.01	38.37	18.1	12.7	47.8	37.8	54.4
3 Rougher Conc 3	5.45	4.73	15.6	5.05	4.1	22.3	5.6
4 Rougher Conc 4	3.14	3.18	4.02	1.94	[illegible]	3.8	1.6
5 Rougher Conc 5	3.54	3.77	2.09	0.62	2.3	2.2	1.2
6 Rougher Tail	75.17	0.45	0.28	0.21	5.5	6.3	8.8
Head (calc)	100.0	5.89	3.35	1.39	100.0	100.0	100.0

Calculated Grades and Recoveries

1+2	12.66	37.95	17.34	11.33	81.8	65.5	79.7
1 to 3	18.15	29.41	16.21	9.20	90.8	87.7	86.6
1 to 4	21.29	25.54	14.41	7.98	92.7	91.5	90.1
1 to 5	24.83	22.43	12.65	6.93	94.5	93.7	91.2

Test No. 4

Purpose:	To produce a flotation concentrate from a 10Kg. charge for Au extraction testwork.
Procedure:	As outlined follows.
Feed:	10 kg. of minus 10 mesh AsPy-Po ore.
Grind:	50 minutes at 65% solids in the large ball mill.

Conditions:

	Reagents Added, grams per tonne					Time, minutes			
	A350	R238	$CuSO_4$	M2030	MIBC	Grind	Cond.	Froth	pH
Grind	-	-	-	-	-	50	-	-	-
Rougher Flot.	25	10	-	-	15	-	1	5	-
Condition	-	-	250	-	-	-	3	-	-
Rougher Flot.	20	5	-	-	5	-	1	3	-
	25	-	-	10	-	-	1	4	-

Metallurgical Results

Product	Weight %	Assays %				% Distribution			
		Au	As	Fe	S	Au	As	Fe	S
1. Ro. Flot. Conc.	16.49	33.47	9.38	26.3	15.8	85.5	86.5	38.9	69.0
2. Ro. Flot. Tail	83.51	1.12	0.29	8.15	1.40	14.5	13.5	61.1	31.0
Head (Calc)	100.00	6.45	1.79	11.1	3.77	100.0	100.0	100.0	100.0

Screen Analysis

Mesh Size (Tyler)	% Retained		% Passing Cumulative
	Individual	Cumulative	
+ 65	0.3	0.3	99.7
100	2.1	2.4	97.6
150	8.1	10.5	89.5
200	15.1	25.6	74.4
270	13.7	39.3	60.7
400	12.3	51.6	48.4
- 400	48.4	100.0	-
Total	100.0	-	-

Test No. 4A

Purpose: To regrind the rougher tailing and float two scavenger concentrates.

Procedure: As indicated below

Feed: 7500 g. of rougher tailing from test No. 4

Grind: 10 minutes at 65% solids in the 10 kg. ball mill.

Conditions:

Stage	Reagents Added, grams per tonne			Time, minutes			pH
	AX350	AP208	MIBC	Grind	Cond.	Froth	
Grind	-	-	-	10	-	-	-
Scavenger No. 1	50	20	10	-	1	5	-
Scavenger No. 2	50	10	5	-	1	5	-

Metallurgical Results

Product	Weight,% Ind.	Weight,% Overall	Assays, g/t, Au	% Distribution Ind. Au	% Distribution Overall Au
1. Scav. Conc. No. 1	4.02	3.36	13.67	38.9	5.6
2. Scav. Conc. No. 2	3.96	3.31	3.87	10.9	1.6
3. Scav. Tailing	92.02	76.85	0.77	50.2	7.3
Head (Calc.)	100.0	83.51	1.41	100.0	14.5

Calculated Grades and Recoveries

Scav. Concs. 1+2	7.98	6.67	8.81	49.8	7.2

Screen Analysis - Scav. Tailing

Mesh Size (Tyler)	% Retained Individual	% Retained Cumulative	% Passing Cumulative
+ 65	0.1	0.1	99.9
100	0.5	0.6	99.4
150	4.3	4.9	95.1
200	9.6	14.5	85.5
270	13.2	27.7	72.3
400	13.2	40.9	59.1
- 400	59.1	100.0	-
Total	100.0	-	-

Overall Metallurgical Results

Product	Weight %	Assays %			% Distribution
		Au	Fe	As	Au
1. Ro. Flot. Conc.	16.49	33.47	-	-	82.4
2. Scav. Conc	6.67	8.81	-	-	8.8
3. Rougher Tailing	76.84	0.77	8.81	0.28	8.8
Head (Calc)	100.00	6.70	-	-	-

Calculated Grades and Recoveries

Products 1 + 2	23.16	26.37	-	-	91.2

Test No. 4A

Test No. 5

Purpose: To repeat test No. 4, but combine all flotation concentrates

Procedure: As indicated below

Feed: 10 kg. of minus 10 mesh AsPy-Po ore.

Grind: 60 minutes at 65% solids in the large ball mill.

Conditions:

	Reagents Added, grams per tonne					Time, minutes			pH
	A350	R208	$CuSO_4$	M2030	MIBC	Grind	Cond.	Froth	
Grind	-	-	-	-	-	60	-	-	-
Rougher Flot.	50	20	-	-	15	-	1	5	-
	50	10	-	-	5	-	1	5	-
Condition	-	-	250	-	-	-	3	-	-
Ro.Flot (Cont)	50	10	-	-	5	-	1	5	-
	25	-	-	10	-	-	1	5	-
	25	5	-	-	5	-	1	3	-

Metallurgical Results

Product	Weight %	Assays %				% Distribution			
		Au	As	S	Fe	Au	As	S	Fe
1. Ro. Flotation Conc.	24.82	22.47	6.44	11.0	23.9	91.7	90.6	87.5	49.2
2. Ro. Flotation Tail	75.18	0.67	0.22	0.52	8.16	8.3	9.4	12.5	50.8
Head (Calc)	100.00	6.08	1.76	3.12	12.1	100.0	100.0	100.0	100.0

Test No. 5

Purpose: To evaluate Au recovery from flotation Conc. by Cyanidation.

Procedure: Sample was preaerated at a pH of 10.5-11.0 maintained with CaO for a 8 hr. period.

Feed: 250 g Ro. Flot. Conc. from Test 4

Solution Volume: 500 mL Pulp Density 33 % solids

Solution Composition: 2.0 gpL NaCN

pH Range: 10.5 - 11.0 with $Ca(OH)_2$

Grind: None

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1	-	2.04	-	1.55	-	-	-	-	11.1-9.8
1-2	-	0.57	-	0.43	-	-	-	-	11.0-10.3
2-4	-	0.20	-	0.15	-	-	-	-	11.2-10.3
4-7	-	0.20	-	0.15	-	-	-	-	11.2-10.1
0-.5	1.06	0.50	1.0	0.38	0.33	-	0.67	-	10.7-9.7
0.5-1	0.70	0.30	0.67	0.23	0.75	-	0.25	-	10.9-10.4
1-1.5	0.26	0.20	0.25	0.15	0.83	-	0.17	-	11.0-10.7
1.5-2.5	0.18	-	0.17	-	0.80	-	0.20	-	10.7-10.7
2.5-4	0.21	0.19	0.20	0.14	0.60	-	0.40	-	11.0-10.8
4-6.5	0.42	-	0.40	-	0.68	-	0.32	-	10.8-10.7
6.5-9.0	0.34	-	0.32	-	0.70	-	0.30	-	10.7-10.6
9-16	0.32	0.20	0.30	0.14	0.93	-	0.07	-	11.4-11.3
16-24	0.07	-	0.07	-	0.37	0.01	0.63	3.31	11.3-10.2
Total	3.56	4.40	3.38	3.32	0.37	0.01	3.01	3.31	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 12.04 kg/t CaO: 13.24 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. 24 hr. Preg Solution	548 mL	3.63	25.2
2. 24 hr. Wash Solution	565 mL	0.56	4.0
3. 24 hr. Residue	239.4 g	23.33	70.8
Head (Calc.)	239.4 g	33.0	100.0

Test No. 6

Purpose:	To evaluate Au recovery from feed sample by direct cyanidation.
Procedure:	As per test 5, but cyanide for 48 hours.
Feed:	500 g -10 mesh
Solution Volume:	1000 mL Pulp Density 33 % solids
Solution Composition:	2.0 gpL NaCN
pH Range:	10.5 - 11.0 with $Ca(OH)_2$
Grind:	23 min/1kg. at 50% solids.

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1	-	2.71	-	2.06	-	-	-	-	11.0-9.7
1-2	-	0.60	-	0.47	-	-	-	-	11.3-10.2
2-4	-	0.60	-	0.47	-	-	-	-	11.6-10.4
4-8	-	0.25	-	0.19	-	-	-	-	11.4-8.7
0-1	2.13	0.60	2.0	0.47	1.15	-	0.85	-	11.1-10.2
1-2.5	0.89	0.50	0.85	0.38	1.40	-	0.60	-	11.2-10.8
2.5-7.0	0.63	0.14	0.60	0.11	1.15	-	0.85	-	11.0-10.7
7.0-10	0.89	-	0.85	-	1.65	-	0.35	-	10.7-10.6
10-16	0.37	0.20	0.35	0.15	1.20	-	0.80	-	11.3-10.8
16-24	0.84	-	0.8	-	1.40	-	0.60	-	10.8-10.3
24-32	0.63	0.30	0.6	0.23	1.85	-	0.15	-	11.3-10.8
32-48	0.16	-	0.15	-	1.45	0	0.55	4.53	10.8-10.6
Total	6.54	5.9	6.2	4.53	1.45	0	4.75	4.53	-

Reagent Consumption (kg/t of cyanide feed)   NaCN: 9.50 kg/t   CaO: 9.06 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. 48 hr. Preg. Solution	980 mL	0.84	28.8
2. 48 hr. Wash	1000 mL	0.12	4.2
3. 48 hr. Residue	500 g	3.83	67.0
Head (Calc.)	500 g	5.71	100.0

Test No. 7

Purpose: To evaluate Au recovery in relation to grind time.

Procedure: Sample was preaerated for 8 h at pH 10.5-11.0 with CaO. Cyn. was carried out in 2.5 L bottles on rolls with NaCN maintained at 2.0 g/L and pH at 10.5-11.0 with CaO for 2 x 24 h stages. Between stages, pulp was filtered, washed and repulped.

Feed: 250 g Combined Ro. Conc. of test 4

Solution Volume: 500 mL Pulp Density 33 % solids

Solution Composition: 2.0 gpL NaCN

pH Range: 10.5 - 11.0 with $Ca(OH)_2$

Grind: 15 min./500 g at 50% solids in lab pebble mill.

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1.5	-	1.5	-	1.14	-	-	-	-	10.6-8.2
1.5-3.0	-	1.2	-	0.91	-	-	-	-	10.5-9.0
3.0-6.0	-	0.5	-	0.38	-	-	-	-	10.8-9.2
6.0-8.0	-	0.5	-	0.38	-	-	-	-	11.1-9.3
0-1.5	-	0.3	1.0	0.23	0.40	-	0.60	-	10.5-10.0
1.5-5.0	0.63	0.4	0.60	0.30	0.75	-	0.25	-	11.2-10.4
5.0-7.0	0.26	0.3	0.25	0.23	0.85	-	0.15	-	11.2-10.7
7.0-24	0.16	0.2	0.15	0.15	0.92	0	0.08	-	11.2-10.3
Total	1.05	4.9	2.00	3.72	0.92	0	1.08	3.72	-
0.3.5	1.05	0.30	1.00	0.23	1.00	-	0.00	-	11.7-10.9
3.5-19	-	-	-	-	0.73	-	0.27	-	10.9-10.5
19-24	0.28	0.10	0.27	0.08	0.67	-	0.33	0.31	10.9-10.8
Total	1.33	0.40	1.27	0.31	0.67	0	0.60	0.31	-
Total	2.38	5.30	3.27	4.03	1.59	0	1.68	4.03	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 7.04 kg/t CaO: 16.88 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. 24 hr. Preg & Wash	1150 mL	2.32	39.3
2. 48 Preg & Wash	1060 mL	0.11	1.5
3. 48 hr. Residue	238.8 g	16.84	59.2
Head (Calc.)	238.8 g	28.4	100.0

AUTOCLAVE LEACH TEST NO. 8

Purpose:	To conduct a pressure oxidation and cyanidation test
Sample:	Rougher float Conc. from Test 4
Grind:	15 min/250g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 16.7% solids in the autoclave vessel. The pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 250

Solution Volume (mL): 1250

Pulp Density 16.7 % Solids

Temperature (°C): 200

Oxygen Over Pressure (psig): 100

Steam Pressure (at temperature) (psig): 225

Total Pressure (psig): 325

Time (at temperature) (min): 90

Autoclave Leach Data:

Time (Minutes)	Temperature °C	Total Pressure psig	O_2 FlowRate, ml/min	O_2%
10	200	330	48	59
15	205	330	84	56
17	205	330	84	49
30	200	330	66	58
36	200	330	128	73
46	200	330	66	85
68	200	330	48	91
80	200	330	48	93
85	200	330	37	95
90	200 End of Test			

Results:

Final pH: 0.8

Final EMF: 520 mv

Free Acid (g/L H_2SO_4): 45.0

Pregnant Solution - Volume (mL): 1110

Observations:

Solution Colour:	Green
Residue Colour:	Red

General Observations: Elemental Sulphur present on walls and lid of autoclave.

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

Test No. 8

Purpose: To evaluate Au recovery by cyanidation.

Procedure: The autoclave residue was pulped with water at 30% solids in a 2.5 L bottle and conditioned with lime until the pH stabilized at approximately 11. Preattritioned carbon and NaCN were added and the CIL test was conducted over 24 hours maintaining 2 g/L NaCN. The pulp was screened, to remove carbon, then filtered and washed.

Feed: 250 g Autoclave Residue

Solution Volume: 583 mL Pulp Density 30 % solids

Solution Composition: 2.0 gpL NaCN

pH Range: 10.5 - 11.0

Carbon g/L: 10

Cyanide Data:

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Cyanidation									
0-1	1.17	14.8	1.11	11.35	0.73	-	0.38	-	11.0-10.8
1-4	0.40	-	0.38	-	0.95	-	0.16	-	10.810.7
4-7	0.17	-	0.16	-	1.03	-	0.07	-	10.7-10.5
7-24	0.07	1.00	0.07	0.76	1.12	0	0.01	-	11.1-10.6
Total	1.81	15.8	1.72	12.01	1.12	0	0.60	-	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 2.40 kg/t CaO: 48.0 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. Carbon	7.2 g	826.7*	95.8
2. Barren Solution + Wash	1210 mL	0.02	0.1
3. Residue	240.0 g	1.07	4.1
Head (Calc.)	250.0 g	24.8	100.0

* Reassay carbon - 905 g/t Au

Additional Assays

Autoclave Solution, mg/L

Fe^{3+} - 5010

Fe^{2+} - <10

As - 940

Size Analysis - 24 hr. Cyn. Res.

Particle Size	% Retained Individual	% Retained Cumulative	% Passing Cumulative
+36.4 µm	2.7	2.7	97.3
28.2	2.1	4.8	95.2
19.7	5.2	10.0	90.0
13.5	6.3	16.3	83.7
10.5	4.8	21.1	78.9
-10.5µm	78.9	100.0	-
Total	100.0	-	-

AUTOCLAVE LEACH TEST NO. 9

Purpose:	To conduct a pressure oxidation and cyanidation test
Sample:	Combined Rougher float Conc. from Test 4
Grind:	25 min/500 g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 16.7% solids in the autoclave vessel. The pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 250
Solution Volume (mL): 1250
Pulp Density 16.7 % Solids
Temperature (^oC): 200
Oxygen Over Pressure (psig): 100
Steam Pressure (at temperature) (psig): 226
Total Pressure (psig): 326
Time (at temperature) (min): 90
Time (heat-up) (min): 50

Autoclave Leach Data:

Time (Minutes)	Temp ^oC	Pressure psig	Off-Gas FlowRate, mL/min	O2 %	Remarks
0	25	30	-	-	Heat up
15	55	30	-	-	Heat up
30	125	70	-	-	Heat up
45	175	175	-	-	Heat up
0	200	320	66	58	Oxidation
10	200	325	106	86	
20	200	325	84	90	
30	200	325	106	94	
40	195	325	66	96	
50	200	325	84	96	
60	200	325	106	96	
70	200	325	84	97	
80	200	325	106	98	
90	200	325	84	98	

Results:

Final pH: 0.5

Final EMF: 600 mv

Free Acid (g/L H_2SO_4): 29.0

Pregnant Solution - Volume (mL): 900

Observations:

Solution Colour:	Yellow-Green
Residue Colour:	Orange

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

Test No. 9

Procedure: The autoclave residue was pulped with water at 30% solids in a 2.5 L bottle and conditioned with lime until the pH stabilized at approximately 11. Preattritioned carbon and NaCN were added and the CIL test was conducted over 24 hours maintaining 2 g/L NaCN (Carbon was changed after 6 and 16 hours). The pulp was screened, to remove carbon, then filtered and washed.

Feed: 205 g Autoclave Residue

Solution Volume: 478

Pulp Density : 30 % solids

Solution Composition: 2.0 gpL NaCN

pH Range: 10.5 - 11.0

Carbon g/L: 10

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1.5	1.01	10.0	0.96	7.60	0.74	-	0.22	-	11.8-11.1
1.5-6.0	0.23	-	0.22	-	0.27	-	0.69	-	11.1-10.5
6.0-16	0.73	0.40	0.69	0.30	0.66	-	0.30	-	11.3-10.8
16.0-18	0.32	-	0.30	-	0.17	-	0.17	-	10.8-10.5
18-24	0.18	0.30	0.17	0.23	0.96	0	0.0	8.13	11.2-10.5
Total	2.47	10.70	2.34	8.13	0.96	0	1.38	8.13	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 6.73 kg/t CaO: 39.7 kg/t

Metallurgical Results

Product	Amount	Assays %,g/t,mg/L				% Distribution
		Au	$S^=$	Cu	Zn	Au
1. 6h Carbon	5.3 g	833	-	-	-	89.8
2. 16h Carbon	4.9 g	37	-	-	-	3.7
3. 24h Carbon	4.8 g	3.0	-	-	-	0.3
4. 24h Barren	1440 mL	0.002	-	4.96	7.05	<0.1
5. 24h Residue	191.6 g	1.58	0.25	-	-	6.1
Head (Calc)	205.0 g	24.0	-	-	-	100.0

Additional Assays

Autoclave Solution, mg/L

Fe^{3+} - 2375

Fe^{2+} - 125

AUTOCLAVE LEACH TEST NO. 10

Purpose:	To conduct a pressure oxidation and cyanidation test
Sample:	Combined Rougher float Conc. from Test 4
Grind:	25 min/500 g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 16.7% solids in the autoclave vessel. The pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 250

Solution Volume (mL): 1250

Pulp Density 16.7 % Solids

Temperature (°C): 200

Oxygen Over Pressure (psig): 100

Steam Pressure (at temperature) (psig): 226

Total Pressure (psig): 326

Time (at temperature) (min): 120

Time (heat-up (min): 55

Autoclave Leach Data:

Time (Minutes)	Temp °C	Pressure psig	Off-Gas FlowRate, mL/min	O2 %	Remarks
0	25	25	-	-	Heat up
15	55	35	-	-	
30	150	95	-	-	
45	180	170	-	-	
0	200	330	84	62	Oxidation
10	200	325	66	89	
20	200	325	66	86	
30	200	325	84	94	
40	200	325	84	95	
50	200	325	106	97	
60	200	340	84	97	
70	200	325	66	98	
80	200	325	84	87	
90	200	325	48	98	
100	200	325	66	97	
110	200	325	66	97	
120	200	325	84	98	

Results:

Final pH 0.5

Final EMF 610 mv

Free Acid (g/L H_2SO_4): 32.0

Pregnant Solution - Volume (mL): 1075

Observations:

Solution Colour: Green

Residue Colour: Orange

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

TEST NO. 10

CYANIDATION

Procedure:	The autoclave residue was pulped with water at 30% solids in a 2.5 L bottle and conditioned with lime until the pH stabilized at approximately 11. Preattritioned carbon and NaCN were added and the CIL test was conducted over 24 hours maintaining 2 g/L NaCN Carbon was changed after 6 and 12 hours. The pulp was screened, to remove carbon, then filtered and washed.
Feed:	250 g Autoclave Residue
Solution Volume:	583 mL
Pulp Density:	30% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with Ca(OH)2
Carbon:	10 g/L

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0.1.5	1.22	11.0	1.17	8.36	0.64	-	0.53	-	11.0-10.5
1.5-3	0.56	0.5	0.53	0.36	1.11	-	0.08	-	11.3-10.9
3-6	0.08	-	0.08	-	0.76	-	0.41	-	10.9-10.7
6-12	0.43	0.4	0.41	0.30	0.32	-	0.85	-	11.3-10.7
12-24	0.89	0.3	0.85	0.23	1.17	0.04	0	9.21	11.3-10.4
Total	3.18	12.2	3.04	9.25	1.17	0.04	1.87	9.21	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 7.48 kg/t CaO:36.8 kg/t

Metallurgical Results

Product	Amount	Assays %,mg/L,g/t				% Distribution
		Au	As	S$^{=}$	Fe	Au
1. 6h Carbon	6.5g	832.00	-	-	-	96.1
2. 12h Carbon	6.1g	10.00	-	-	-	1.1
3. 24h Carbon	6.0g	1.30	-	-	-	0.1
4. 24h Barren Sol'n + Wash	1260 mL	<0.002	-	-	-	0.1
5. 24h Residue	233.0 g	0.63	7.89	0.056	20.5	2.6
Head (Calc)	250.0 g	22.5	-	-	-	100.0

Additional Assays

Autoclave Solution, mg/L

Fe^{3+} - 2416

Fe^{2+} - 34.2

As - 267

TEST NO. 10A (1)

Purpose: To evaluate the stability of arsenic compounds in the CIL residue

Procedure: 50 g of CIL residue was pulped with deionized H_2O, filtered and washed with 2 displacement washes. The filtercake was repulped with DI H_2O to an 800 g slurry weight. The pH was adjusted, with a 10% H_2SO_4 solution, to pH = 8, and then slurry weight was increased to 1000 g. The slurry was agitated and settled daily with pH being adjusted as necessary. At the end of 180 h test, 100 mL of solution was filtered through a 0.45 µm millipore and assayed.

Feed: Test No. 10 - CIL Residue

Flowsheet: See attached flowsheet

Conditions:

Time Hours	pH	10% H_2SO_4 required mL
0-18	8.0 - 9.1	2.4
18 - 25	7.8 - 8.3	1.0
25 - 40	8.3 - 8.7	-
40 - 46	7.6 - 8.5	0.6
46 - 48	7.7 - 7.8	0.5
48 - 64	7.8 - 8.0	-
64 - 72	8.0 - 8.1	-
72 - 88	8.1 - 8.3	-
88 - 96	7.5 - 7.5	0.3
96 - 160	7.5 - 8.0	-
160 - 180	8.0 - 8.0	-

Results: See Tables.

Arsenic Compound Stability Data

1. Conditions

Feed	Test	Neutralization		Test	H2SO4 Req't
		pH	CaO, g/L*	pH	kg/t**
CIL Residue T-10	10-A1	11	-	8	10.7
	10-A2	11	-	8	5.6
Autoclave Liquor T-10	10-B1	5	22.7		
		11	2.0	8	34.7
	10-B2	5	17.7		
		11	7.6	8	35.8

* Autoclave Liquor
** Neutralized Residue

2. Results

Test	Feed Assay %, mg/L		180h Solution Analysis, mg/L		Solids g/L	Residue Assays %	
	Fe	As	Fe	As		Fe	As
10-A1	20.50%	7.89%	1.55	7.9	45	23.1	8.51
10-A2	20.50%	7.89%	1.62	9.97	45	23.3	8.51
10-B1	2450mg/L	267mg/L	0.05	0.41	30	2.38	0.27
10-B2	2450mg/L	267mg/L	0.17	0.14	24	2.71	0.31

TEST NO. 10A (2)

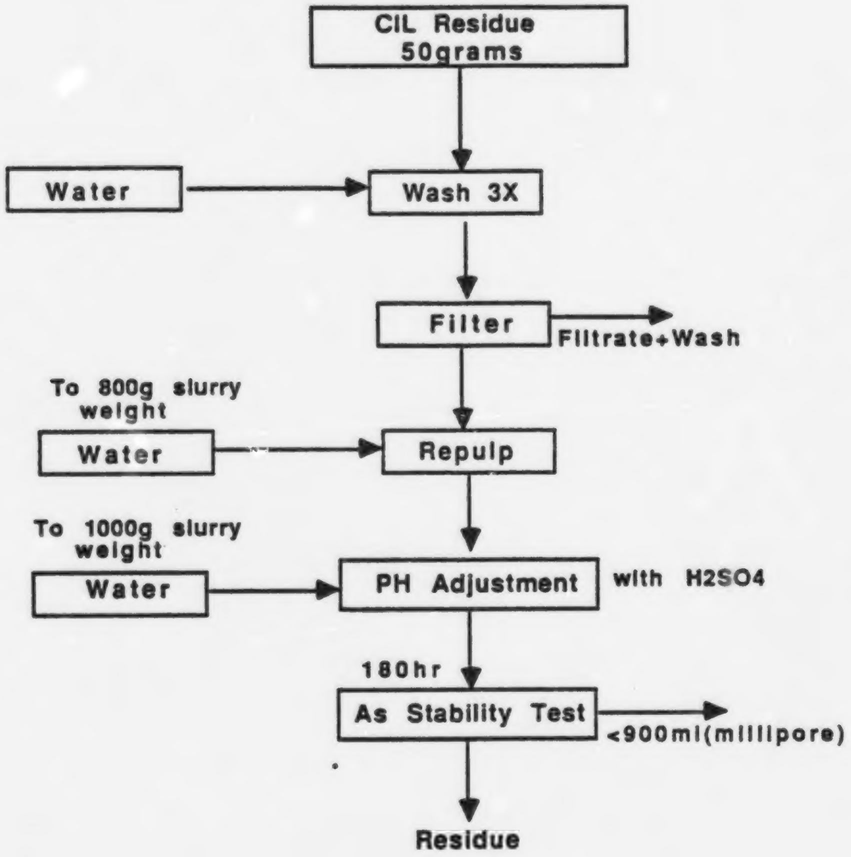
Purpose:	To repeat conditions of Test 10 A (1)
Procedure:	As per Test 10A (1)
Feed:	Test No. 10 - CIL Residue
Flowsheet:	See attached flowsheet

Conditions:

Time Hours	pH	10% H_2SO_4 required mL
0-18	8.0 - 8.8	1.8
18 - 25	7.5 - 8.0	0.5
25 - 40	8.8 - 8.0	-
40 - 46	8.0 - 8.2	-
46 - 48	7.8 - 7.8	0.2
48 - 64	7.8 - 8.0	-
64 - 72	8.0 - 8.0	-
72 - 88	8.0 - 8.0	-
88 - 96	8.0 - 7.5	-
96 - 160	7.5 - 8.0	-
160 - 180	8.0 - 8.1	-

Results: See Tables.

Flowsheet


CIL Residue
50grams
Water
Wash 3X
Filter
Filtrate+Wash
To 800g slurry weight
Water
Repulp
To 1000g slurry weight
Water
PH Adjustment
with H2SO4
180hr
As Stability Test
<900ml(millipore)
Residue

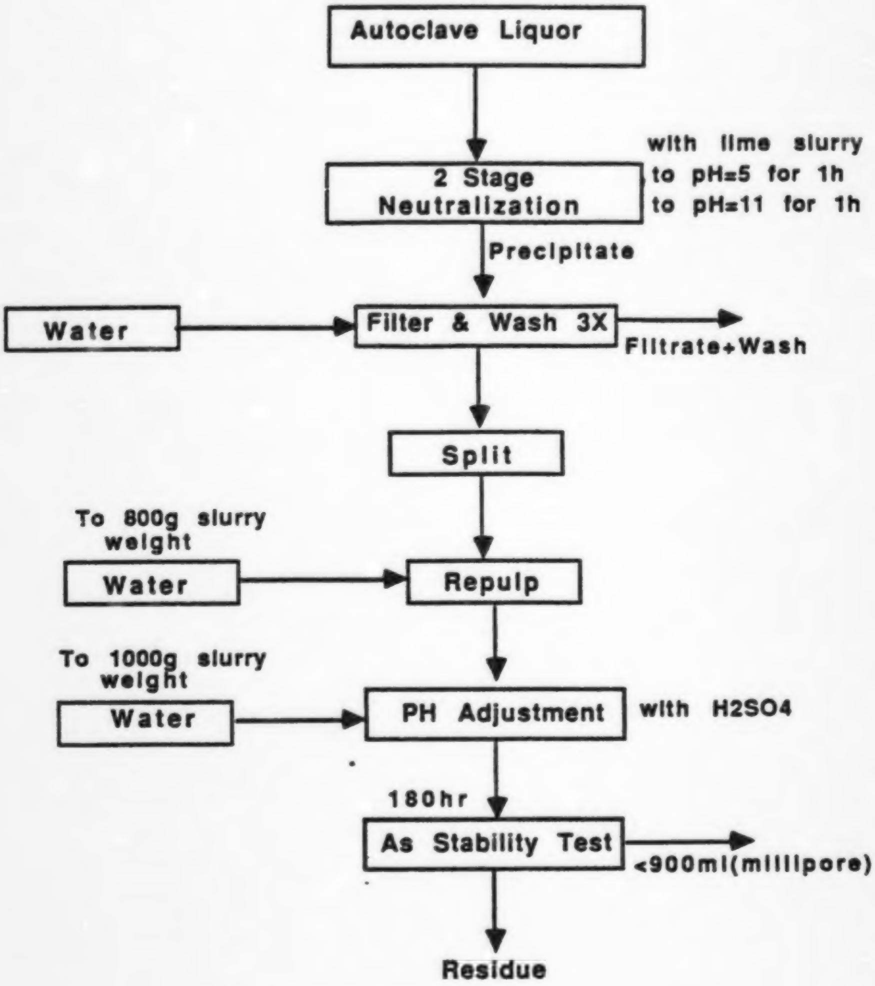
TEST NO. 10B (1)

Purpose:	To evaluate the stability of arsenic compounds in the neutralized autoclave liquor
Procedure:	300 mL autoclave liquor was neutralized, with a 20% $Ca(OH)_2$ slurry, at pH = 5 for 1 h and then at pH - 11 for another hour. The solution was filtered, washed 3 times by repulping and then ~ 50 g of precipitate was repulped to an 800 g slurry weight with DI H_2O. pH was adjusted to 8 with a 10% H_2SO_4 solution, and slurry weight was increased to 1000 g. The slurry was agitated and settled daily with pH being adjusted, as necessary. At end of 180 h test, 100 mL of solution was filtered through a 0.45 µm millipore and assayed.
Feed:	Test No. 10 - Autoclave Liquor
Flowsheet:	See attached
Conditions:	

Time Hours	pH	Reagents Required, mL 10% H_2SO_4	 20% $Ca(OH)_2$
1 - neutr.	5	-	45
2 - neutr.	11	-	4
0-18	8.1 - 9.2	6	-
18 -25	8.0 -7.8	1.5	-
25-40	7.8 - 7.6	-	-
40-46	7.6 - 9.0	-	-
46-48	8.0 - 7.6	1.3	-
48-64	7.6 - 8.5	-	-
64-72	7.7 - 8.5	0.6	-
72-88	8.0 - 8.7	0.5	-
88-96	8.0 - 7.5	0.5	-
96-160	7.5 - 7.8	-	-
160-180	7.8 - 8.2	-	-

Results: See Tables.

Flowsheet


Autoclave Liquor
2 Stage Neutralization
with lime slurry
to pH=5 for 1h
to pH=11 for 1h
Precipitate
Water
Filter & Wash 3X
Filtrate+Wash
Split
To 800g slurry weight
Water
Repulp
To 1000g slurry weight
Water
PH Adjustment
with H2SO4
180hr
As Stability Test
<900ml(millipore)
Residue

TEST NO. 10B (2)

Purpose:	To repeat conditions of Test 10B (1)
Procedure:	As for Test 10B (1)
Feed:	Test No. 10 - Autoclave Liquor
Conditions:	

Time Hours	pH	Reagents Required, mL 10% H_2SO_4	 20% $Ca(OH)_2$
1 - neutr.	5	-	35
2 - neutr.	11	-	15
0-18	7.6 - 8.8	3	-
18-25	7.8 - 8.0	3.5	-
25-40	8.0 - 8.0	-	-
40-46	8.0 - 9.1	-	-
46-48	8.0 - 7.8	1.5	-
48-64	7.8 - 8.3	-	-
64-72	7.9 - 7.6	0.6	-
72-88	7.6 - 8.0	-	-
88-96	8.0 - 7.5	-	-
96-160	7.5 - 7.9	-	-
160-180	7.9 - 8.2	-	-

Results: See Tables.

TEST NO. 11

CYANIDATION

Purpose: As per test 7

Procedure: As per test 7

Feed: 250g combined Ro Conc. of test 4

Solution Volume: 500 mL Pulp Density 33% Solids

Solution Composition: 2.0 g/L NaCN

pH Range: 10.5-11 with $Ca(OH)_2$

Grind: 4 x 30 min/250g @ 50% solids in lab pebble mill

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1.5	-	1.7	-	1.30	-	-	-	-	10.6-8.0
1.5-3.0	-	1.7	-	1.30	-	-	-	-	10.6-9.3
3.0-6.0	-	0.5	-	0.38	-	-	-	-	10.7-9.4
6.0-8.0	-	0.5	-	0.38	-	-	-	-	11.0-9.4
0-1.5	1.06	0.3	1.0	0.23	0.20	-	0.80	-	10.5-10.2
1.5-5.0	0.84	0.4	0.80	0.30	0.58	-	0.42	-	11.1-10.4
5.0-7.0	0.44	0.3	0.42	0.23	0.85	-	0.15	-	11.2-10.7
7.0-24	0.16	0.2	0.15	0.15	0.73	0	0.27	-	11.1-10.3
Total	2.50	5.6	2.37	4.27	0.73	0	1.64	4.27	-
0-3.5	1.05	0.30	1.00	0.23	1.00	-	0.00	-	11.7-10.8
3.5-19	-	-	-	-	0.40	-	0.60	-	10.8-10.3
19-24	0.63	0.20	0.60	0.15	0.55	0.01	0.45	0.37	11.0-10.7
Total	1.68	0.50	1.60	0.38	0.55	0.01	1.05	0.37	-
Total	4.18	6.10	3.97	4.65	1.28	0.01	2.69	4.64	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 11.06 kg/t CaO: 19.07 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. 24 h Preg + Wash	1220 mL	2.03	39.6
2. 48 h Preg + Wash	900 mL	0.085	1.2
3. 48h Residue	243.3 g	15.19	59.2
Head (Calc.)	243.3 g	25.7	100.0

Size Analysis - 48 h Residue

Particle Size	% Retained		% Passing Cumulative
	Individual	Cumulative	
+30.1µm	0.1	0.1	99.9
21.0	0.8	0.9	99.1
14.5	3.1	4.0	96.0
11.2	4.0	8.0	92.0
-11.2µm	92.0	100.0	-
Total	100.0	-	-

AUTOCLAVE LEACH TEST NO. 12

Purpose:	To conduct a pressure oxidation test using HNO_3.
Sample:	125 g Rougher Conc. from Test 5
Grind:	12.5 min/250g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 9.1% solids in the autoclave vessel. HNO_3 was added and the pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 125

Solution Volume (mL): 1250

Pulp Density 9.1 % Solids

Temperature (°C): 80

Total Pressure (psig): 100

Time (at temperature) (min): 60

Time (heat-up (min): 10

HNO_3 (g/L): 62.5

Autoclave Leach Data:

Time (Minutes)	Temp °C	Pressure psig	Off-Gas FlowRate, mL/min	O2 %	Remarks
0	25	25	No Off-Flow		Heat up
10	80	250			
0	80	250			Oxidation
5	80	260			
10	90	280			
15	95	260			
20	95	240			
25	95	220			
30	90	200			
35	90	180			
40	90	165			
45	85	150			
50	85	140			
55	80	125			
60	80	115			

Results:

Final pH 1.35

Final EMF 700 mv

Free Acid (g/L H_2SO_4): 4.7

Observations:

Solution Colour: Brownish Yellow

Residue Colour: Yellow

General Observations: Elemental Sulfur formed

Wgt = 1.85 g

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

TEST NO. 12

CYANIDATION

Procedure:	The autoclave residue was pulped with water at 30% solids in a 2.5 L bottle and conditioned with lime until the pH stabilized at approximately 11. Preattritioned carbon and NaCN were added and the CIL test was conducted over 24 hours maintaining 2 g/L NaCN. The pulp was screened, to remove carbon, then filtered and washed. (Sulphur beads were not cyanided).
Feed:	125 g Autoclave Residue
Solution Volume:	291 mL
Pulp Density:	30% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with Ca(OH)2
Carbon:	10 g/L

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-2	0.61	9.5	0.58	7.22	0.33	-	0.25	-	11.2-10.5
2-6	0.26	0.5	0.25	0.38	0.25	-	0.33	-	11.3-10.8
6-19	0.35	-	0.33	-	0.18	-	0.40	-	10.8-10.7
19-24	0.42	-	0.40	-	0.22	0	0.36	-	10.7-10.6
Total	1.64	10.0	1.56	7.6	0.22	0	1.34	7.6	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 10.72 kg/t CaO:60.8 kg/t

Metallurgical Results

Product	Amount	Assays %			% Distribution	
		Au	S=	S°	Ind. Au	O'll Au
1. Autoclave Leach Solution	1230 mL	0.60	-	-	36.6	-
2. 6 h Carbon*	2.9 g	235	-	-	33.8	-
3. 24 h Carbon	3.0 g	26.0	-	-	3.9	-
4. 24 h Barren	1020 mL	<0.002	-	-	0.1	-
5. 24 h Residue	112.3 g	0.60	0.008	-	3.3	-
6. S Beads	1.85 g	244	-	68.8	22.3	-
Autoclave Feed (calc)	125.0 g	16.1	-	-	100.0	-

Test No. 12

TEST NO. 13

Purpose:	To investigate the effect of roasting the flotation concentrate on the subsequent cyanide leaching of gold.
Procedure:	Preheat a cast-iron tray in a muffle to 580° C. Add a 273 gram charge of concentrate and continue heating at 580° C for 2.5 hours. Rabble almost continually during the first hour. After the roasting period, remove tray from muffle and air-cool to room temperature.
Feed:	273 grams of Rougher Flotation Concentrate from test No. 5.

Data:

Time	Elapsed Time	Charge Temp.°C	Muffle Temp.°C	Remarks
1:30	0	20	580	
1:31	1	300	-	Burning, blue flame
1:35	5	425	-	Burning, Fumes
1:40	10	485	-	Glowing, Fumes
1:45	15	520	-	Glowing, Fumes
1:50	20	565	-	Glowing, Fumes
2:00	30	578	-	Glowing, Fumes
2:05	35	600	-	Glowing, Fumes
2:10	40	590	-	Faint glowing, faint fumes
2:15	45	585	-	glowing & fumes almost ceased
2:20	50	565	-	glowing & fumes ceased
2:30	60	550	-	
2:40	70	515	-	
2:50	80	500	-	close door
3:00	90	510	-	
3:15	105	535	-	
3:30	120	575	-	
4:00	150	580	-	
4:30	180	580	-	charge out

Observations: Concentrate started to burn immediately, with blue flame and giving off white fumes. Burning continued for 45 minutes. During burning, the charge temperature climbed to 600° C and after burning ceased, temperature dropped to 500° C. Temperature of 580° C could only be maintained with door closed.

Calcine: Rusty brown in colour.

Calcine Wt: 225.5 g or 82.6% wt. of feed.

H_2SO_4 Leaching: The calcine was transferred to a one-liter beaker and leached for one hour at 80°C in 500 mL of 10g/L H_2SO_4 solution. The pulp was then filtered and washed, the solution measured and assayed, and the residue prepared for cyanide leaching.

Solution Analysis:

As - 267 mg/L
S - 4400 mg/L
Fe_T-94.2 mg/L

TEST NO. 13

CYANIDATION

Purpose: To investigate the extraction of gold from the calcine by cyanide leaching.

Procedure: The calcine was pulped in a 2.5 L bottle, placed on rolls, and cyanided as described below for a 24h retention time. Pulp was then filtered, washed and products submitted for assay.

Feed: 225 g calcine

Solution Volume: 450 mL Pulp Density 33% Solids

Solution Composition: 2.0 g/L NaCN

pH Range: 10.5 with Ca(OH)2

Carbon: 10 g/L 4.5 g

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1.5	0.95	1.45	0.90	1.10	0.47	-	0.43	-	11.6-9.5
1.5-4.0	0.45	0.50	0.43	0.38	0.90	-	-	-	11.0-10.1
4.0-7.0	-	0.30	-	0.23	0.90	-	-	-	11.1-10.5
7.0-24	-	0.20	-	0.15	0.71	0	0.19	1.86	11.0-10.5
Total	1.40	2.45	1.33	1.86	0.71	0	0.62	1.86	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 2.82 kg/t CaO: 8.47 kg/t

Metallurgical Results

Product	Amount	Assays,%, g/t, mg/L Au	 S=	%Distribution Au
1. 24 Hour Carbon	5.4 g	727.5	-	71.0
2. 24 Hour Barren + Wash	965mL	0.012	-	0.2
3. 24 Hour Residue	219.5 g	7.24	0.023	28.8
Head (Calc.)	219.5 g	25.19	-	100.0

AUTOCLAVE LEACH TEST NO. 14

Purpose:	To repeat Test 12 with higher O_2 pressure
Sample:	125 g Rougher Conc. from Test 5
Grind:	12.5 min/250g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 9.1% solids in the autoclave vessel. HNO_3 was added and the pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 120

Solution Volume (mL): 1200

Pulp Density 9.1 % Solids

Temperature (°C): 80

Total Pressure (psig): 300

Time (at temperature) (min): 60

Time (heat-up) (min): 14

HNO_3 (g/L): 62.5

Autoclave Leach Data:

Time (Minutes)	Temp °C	Pressure psig	Off-Gas FlowRate, mL/min	O2 %	Remarks
0	25	25	No off-flow	-	Heat up
5	30	25	-	-	
10	40	45			
14	80	170			
0	80	210			Oxidation
5	105	310			
10	115	300			
15	120	310			
20	120	310			
25	115	300			
30	110	310			
35	105	300			
40	105	300			
45	100	300			
50	90	300			
55	90	300			
60	90	270			

Results:

Final pH	0.6
Final EMF	+700 mv

Free Acid (g/L H_2SO_4): 27.7

Pregnant Solution - Volume (mL): 1140

Observations:

Solution Colour:	Brownish Yellow
Residue Colour:	Yellow
General Observations:	Temperature was considerably higher than in Test 12 and increased over the melting point of S° (119°C). Several large (d=12mm) sulphur beads formed. Reaction pressure reached 170 psi before O_2 was added.

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The residue was not cyanided.

Additional Assays

Autoclave Solution, mg/L

Fe3+ - 18233

Fe2+ - 167

As - 6480

TEST NO. 15

Purpose: To investigate roasting the concentrate before cyanidation.

Procedure: Add 268 gram charge of concentrate to cold cast-iron tray. Place in muffle furnace at 1100°F. Rabble continuously for 1 hour. Try to maintain maximum temperature below 650° C. Maintain temperature at 590°C for additional hour with occasional rabbling. Air cool.

Feed: 268 grams of Rougher Flotation Concentrate from test No. 5.

Data:

Time	Elapsed Time	Charge Temp.°C	Muffle Temp.°F	Remarks
4:00 p.m.	0	-	1100	start...place tray in furnace set at 4:07 1100°F
	-	250	-	
4:15	15	460	1000	copious white fume beginning at 510°C
4:20	-	525	-	
4:25	-	590	1070	furnace turned down to 1000°F
4:30	30	650	-	glowing charge
4:33	-	665	-	
4:35	-	655	-	furnace set at 1100°F
4:40	-	585	-	charge stopped glowing
4:45	45	570	1070	
4:50	-	571	1070	
4:55	-	570	-	
5:00	60	574	1080	
5:05	-	583	1090	
5:10	-	585	-	
5:15	75	590	1090	
5:20	-	591	-	
5:25	-	591	-	
5:30	90	593	1095	
5:35	-	596	-	
5:40	-	595	-	
5:45	105	595	1100	remove charge from furnace
6:05	-	220	-	

Conditions:

Total roasting time = 90 minutes
1st Stage roast time= 30 minutes
2nd stage roast time= 60 minutes
Maximum temperature= 665° C
Soaking temperature = 590°C

Observations:

Copious white fumes ($As2O3$) were emitted by the sample starting at 510°C and continuing for about 15 minutes. The charge temperature reached a maximum of 665°C, decreased to 570°C and was maintained for 3/4 hour at 585 - 600°C to oxidize any pyrite not oxidized in the initial roast.

The furnace had sufficient capacity to maintain temperature with the muffle 1/2 open to admit oxidizing air. The calcine was brick-red.

Results:

Calcine Weight = 225 g or 84.0 % wt. of feed

Description of Apparatus

Lindberg double door muffle furnace with Thermolyne temperature controller. Lower temperature controller door resting on firebrick and cupels Firebrick spacer between doors.

Cast iron tray angled toward back on firebrick. The sample temperature was measured on a Thermoelectric digital pyrometer.

H_2SO_4 Leaching: The calcine was transferred to a one-liter beaker and leached for one hour at 80°C in 500 mL of 10g/L H_2SO_4 solution. The leached pulp was then filtered and washed, the solution measured and assayed, and the residue prepared for cyanide leaching.

Solution Analysis: Fe_T - 39.7 mg/L

TEST NO. 15

CYANIDATION

Purpose: To investigate Au recovery from a previously acid leached calcine by cyanide leaching.

Procedure: As per test 13

Feed: 225 g Calcine

Solution Volume: 450 mL Pulp Density 33% Solids

Solution Composition: 2.0 g/L NaCN

pH Range: 10.5 with Ca(OH)2

Carbon: 10 g/L

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	Ca(OH)$_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1.5	0.95	2.7	0.90	2.05	0.63	-	0.27	-	10.8-10.1
1.5-4.0	0.28	0.3	0.27	0.23	0.90	-	-	-	11.0-10.4
4.0-7.0	-	0.1	-	0.08	0.79	-	0.11	-	10.7-10.5
7.0-24	0.12	0.2	0.11	0.15	0.76	0	0.14	2.51	11.0-10.7
Total	1.35	3.3	1.28	2.51	0.76	0	0.52	2.51	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 2.42 kg/t CaO:11.66 kg/t

Metallurgical Results

Product	Amount	Assays %, mg/L,g/t			% Distribution
		Au	$S^{=}$	As	Au
1. 24h Carbon	5.6 g	775	-	-	79.0
2. 24 h Barren + Wash	900 mL	0.014	-	-	0.2
3. 24 h Residue	215.2 g	5.31	0.032	2.10	20.8
Head (calc.)	215.2 g	25.54	-	-	100.0

TEST NO. 15A

Purpose:	To determine quantity of gold associated with arsenates.
Sample:	136.8 g residue from test 15
Procedure:	The cyanide residue from Test 15 was leached with NaOH to dissolve As present as ferric arsenate. The leached residue was cyanided to recover gold associated with the arsenates.
Conditions:	1. NaOH Leach The sample was pulped to 20% solids in a 2000 mL glass reactor. NaOH pellets were added to 150 g/L NaOH. The pulp was boiled for 2 hours at 100°C. The pulp was filtered and the cake was washed by repulping. 2. NaCN Leach The sample was pulped with fresh water at 33% solids and cyanided with 10 g/L NaCN for 24 hours.

Metallurgical Results

Product	Amount %	Assays,mg/L Au	% Distribution Au
1. NaOH Solution	500 mL	0.34	2.2
2. NaCN Solution	1155 mL	0.21	31.2
3. Residue	136.8 g	3.80	66.6
Head (Calc)	136.8	5.7	100.0

AUTOCLAVE LEACH TEST NO. 16

Purpose:	To repeat Test 12 with lower HNO_3 Concentration
Sample:	125 g Rougher Conc. from Test 5
Grind:	12.5 min/250g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 9.1% solids in the autoclave vessel. HNO_3 was added and the pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 125
Solution Volume (mL): 1250
Pulp Density 9.1 % Solids
Temperature (°C): 80
Total Pressure (psig): 100
Time (at temperature) (min): 90
Time (heat-up (min): 45 to T = 50°C
HNO_3 (g/L): 20.8

Autoclave Leach Data:

Time (Minutes)	Temp °C	Pressure psig	Off-Gas FlowRate, mL/min	O_2 %	Remarks
0	50	10	No off-flow		Apply 10 psi 0_2
5	65	10			
10	70	10			
15	87	10			
20	95	10			
25	95	10			
30	95	10			
35	90	15			Increase 0_2 pressure
40	90	30			
45	87	50			Set Temp to 85°C
50	87	100			
55	85	100			
60	83	100			
65	85	100			
70					
75					
80					
85					
90					

Results:

Final pH	1.6
Final EMF	+750 mv
Free Acid (g/L H_2SO_4):	6.6

Observations:

Solution Colour:	Green
Residue Colour:	Orange

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

TEST NO. 16

CYANIDATION

Purpose:	As for Test 12	
Procedure:	As per test 12	
Feed:	~125 g Autoclave Discharge	
Solution Volume:	291 mL	Pulp Density 30% Solids
Solution Composition:	2.0 g/L NaCN	
pH Range:	10.5 -11.0 with $Ca(OH)2$	
Carbon:	10 g/L	

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1	0.61	4.25	0.58	-	0.34	-	0.24	-	11.0-11.0
1-23	0.25	-	0.24	-	0.10	-	0.48	-	11.0-10.0
23-28	0.50	0.30	0.48	-	0.37	0.03	0.21	-	11.1-10.9
Total	1.36	4.55	1.30	3.66	0.37	0.03	0.93	3.63	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 7.44 kg/t CaO: 29.0 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. Autoclave Leach Solution	1235	0.52	26.7
2. Carbon	3.1 g	432.0	55.8
3. Barren + Wash Solution	2300 mL	0.007	0.7
4. CN Leach Residue	79.3 g	5.10	16.8
Autoclave Feed (calc)	125.0 g	19.2	100.00

Additional Assays

Autoclave Solution, mg/L
Fe^{3+} - 10780
Fe^{2+} <10
As - 2790

AUTOCLAVE LEACH TEST NO. 17

Purpose:	To repeat test 16, but a lower pulp density
Sample:	75 g Rougher Flot. Conc. from T-5
Grind:	12.5 min/250g at 50% solids in a laboratory pebble mill
Procedure:	The sample was pulped with water at 6.0% solids in the autoclave vessel. HNO_3 was added and the pressure oxidation test was carried out following the conditions listed below. The vessel was then quenched and the pulp filtered.

Leach Conditions:

Feed Weight (g): 75

Solution Volume (mL): 1250

Pulp Density 6.0% Solids

Temperature (^oC): 60

Oxygen Over Pressure (psig): 50-100

Time (at temperature) (min): 60

Time (heat-up) (min.): 30 to T = 60°C

HNO_3 (g/L) - 20.8

Autoclave Leach Data:

Time (Minutes)	Temp °C	Pressure psig	Off-Gas FlowRate, mL/min	O_2 %	Remarks
0	20	-	No off Flow	-	Heat-up
10	25	-		-	Heat-up
15	35	-	No Off Flow	-	Heat-up
20	40	-	-	-	Heat-up
25	50	~10	No Off Flow	-	Heat-up
28	60	~10		-	Heat-up
30	60	10	No Off Flow	-	O_2 Applied
31	65	35	-	-	-
31.5	68	45	-	-	-
31.8	70	58	-	-	-
32.5	73	65	-	-	Shut Off O_2 Supply
33	74	70	-	-	-
34	75	73	-	-	-
35	80	75	-	-	-
36	82	76	-	-	-
37	81	79	-	-	-
38	82	83	-	-	-
39	82	88	-	-	-
40	82	94	-	-	-
41	81	98	-	-	-
42	81	98	-	-	-
45	80	98	-	-	-
46	80	98	-	-	Increasing O_2 30 psi over present pressure
47	84	130	-	-	
48	84	130	-	-	-
49	84	130	-	-	-
50	84	130	-	-	-
51	84	130	-	-	Increasing pressure additional 50 psi over present pressure
52	84	140	-	-	
55	84	140	-	-	temp. stabilizing and falling.
60	82	140	-	-	-
63	80	140	-	-	-
65	78	140	-	-	-
75	75	190	-	-	-
78	73	190	-	-	-
85	73	190	-	-	-
90	70	190	-	-	-

Results:

Final pH 0.9

Final EMF +700 mv

Free Acid (g/L H_2SO_4): 11

Observations:

Solution Colour:	Clear
Residue Colour:	Red

S/L Separation

Procedure: The autoclave leach pulp was filtered to remove the pregnant solution. The unwashed residue was repulped with water and neutralized with lime prior to cyanidation. The lime addition is contained in the cyanide data.

Test No. 17

Procedure: As for Test 12

Feed: 75 g Autoclave Residue

Solution Volume: 175 mL

Pulp Density % solids: 30

Solution Composition: 2.0 gpL NaCN

pH Range: 10.5 - 11.0

Carbon (g/L): 10.0

Reagent Balance:

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
1	-	2.5	-	1.9	-	-	-	-	11.0-10.5
0-1	0.37	0.2	0.35	0.15	0.19	-	0.16	-	11.1-10.7
1-3	0.17	-	0.16	-	0.23	-	0.12	-	10.7-10.3
3-5	0.13	0.2	0.12	0.15	0.28	-	0.07	-	11.3-11.0
5-24	0.07	-	0.07	-	0.17	0.0	0.18	2.2	11.0-10.6
Total	0.74	2.9	0.70	2.2	0.17	0.0	0.53	2.2	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 7.07 kg/t CaO: 29.3 kg/t

Metallurgical Results

Product	Amount	Assays, g/t,mg/L Au	% Distribution Au
1. Autoclave Leach Solution	1225	0.48	30.4
2. Loaded Carbon	1.9 g	600	59.0
3. Barren + Wash Solution	340 mL	0.014	0.3
4. Residue	52.5 g	3.80	10.3
Autoclave Feed (calc)	75.0 g	25.8	100.0

Additional Assays

Autoclave Solution, mg/L
Fe^{3+} - 7625
Fe^{2+} <10
As - 2530

Test No.17A

Purpose: To repeat test No. 5 (A) to produce flotation concentrate for further testwork.

Procedure: As indicated below

Feed: 10 kg. of minus 10 mesh AsPy-Po ore.

Grind: 60 minutes at 65% solids in the large ball mill.

Conditions:

	Reagents Added, grams per tonne					Time, minutes			pH
	A350	R208	$CuSO_4$	M2030	MIBC	Grind	Cond.	Froth	
Grind	-	-	-	-	-	60	-	-	-
Rougher Flot.	50	20	-	-	15	-	1	5	-
	50	10	-	-	5	-	1	3	-
Condition	-	-	250	-	-	-	3	-	-
Rougher Flot.	50	10	-	-	5	-	1	5	-
	25	-	-	10	-	-	1	3	-
	25	5	-	-	5	-	1	3	-

Metallurgical Results

Product	Weight %	Assays %, g/t				% Distribution			
		Au	As	S	Fe	Au	As	S	Fe
1. Ro. Flot. Conc.	21.84	24.4	7.24	12.9	24.1	89.0	89.4	76.8	47.0
2. Ro. Flot. Tail	78.16	0.84	0.24	1.09	7.58	11.0	10.6	23.2	53.0
Head (Calc)	100.00	5.99	1.77	3.67	11.2	100.0	100.0	100.0	100.0

Test No. 18

BIOLOGICAL LEACHING

Innoculator Beaker

Start-up Procedure

TEST NO. 18

Purpose: Test was performed in a 1 L stainless steel beaker which was placed on a hot plate to maintain the temperature at about 35°C. An air sparge was applied to maintain dissolved O_2 concentration above 2 mg/L.

To begin the culture the following reagents were added:

20 mL H_2SO_4 (98%)/liter reactor volume
5 g/L $Fe(SO_4)_3$ - commercial grade
2 g/L Fe SO_4 - reagent grade
9 K media: 2 g/L $(NH4)_2SO_4$
0.5 g/L $MgSO_4$
2.0 g/L $KMPO_4$

~ 150-200 mL pulp of Dickenson Mine concentrate and 95 grams of reground (30 min/250 grams) rougher concentrate of Test 17 were added.

The pulp was allowed to mix for 1 hour, then pH, EMF, T.D.O. were measured and recorded - see attached table.

The slurry conditions were maintained as follows:

T-35-38°C
O2 $\geq$ 2 mg/L
pH - 1.5-2.0
Liquor S.G. - 1.02-1.04
EMF - 350-450 mv

BIOLEACH
DAILY OPERATING DATA
INOCULATOR BEAKER

Date	Oxygen Dissolved (mg/L)	Oxygen Uptake (mg/L/hr)	pH	Temp. (C)	EMF (mV)	Liquor S.G. (g/mL)	Titrated Fe (mg/L) 2+	Titrated Fe (mg/L) 3+	Titrated Fe (mg/L) Total	Other
15/12	7.0	0	1.23	37.1	520	1.049	3001	9799	12800	
16/12	5.5	34	1.35	35.0	530	1.051	2573	10027	12600	
17/12	3.2	45	1.48	52.0	465	1.056	7634	4914	12548	
18/12	4.0	30	1.70	34.8	490	1.057	5128	8739	13867	
	4.8	44	1.60	33.6	500	1.058	-	-	-	
19/12	4.9	34	1.35	35.0	505	1.056	-	-	-	300ml transfered to the reactor

TEST NO. 19

Purpose:	To perform batch biological leaching test.
Procedure:	1 kg of ground flotation concentrate from Test 17 was pulped with water at 10% solids to 10L of total pulp volume. Agitation was provided by a mechanical mixer with a turbine propeller. The reactor was a 15 L baffled container. An air-CO_2 mixture was sparged and evenly distributed through the slurry by a specially designed oxygen diffusor. The reactor was placed in a water bath equipped with an immersion heater. The pH of the slurry was adjusted with H_2SO_4 to 1.8 during 3 days. Initially 100 mL of liquor containing some thiobacillus bacteria from an unknown feed stock was added. Later 300 mL of slurry containing active bacteria (developed on Dickenson Mine concentrate plus concentrate from T-17) was innoculated into the reactor. Baseline data, which includes pH, T, EMF, liquor S.G. was obtained. Liquor samples were removed and analyzed for Fe^{3+}, Fe^{2+}, As. Pulp samples were removed for cyanidation.
Results:	See attached tables.

Bioleach Oxidation Data

Date	Time days	Bio Residue Assays, %			Solids Wgt Remained, g	Liquor Volume, mL	Fe(T) mg/L	% Fe Extraction	% Au Recovery
		Fe	As	S=					
9	1	24.1	7.24	-	1096	9670	4782	17.5	40.9
10	2				-	-	-	-	-
11	3				-	-	-	-	-
12	4				1093	9671	6600	24.2	-
13	5				1091	9671	5400	19.6	-
14	6				1088	9672	5700	21.0	-
15	7				1084	9673	7500	27.8	-
16	8	17.8	6.89	6.76	1024	9692	8098	31.8	54.3
17	9				1024	9692	8597	31.9	-
18	10				1024	9692	10960	40.6	-
19	11				1024	9692	10341	38.4	-
20	12				1024	9692	12140	45.0	-
21	13				1024	9692	15952	59.1	-
22	14				951	9713	13217	56.0	67.8
23	15	6.51	1.62	0.81	888	9732	14755	67.1	84.0
24	16				826	9751	13610	66.7	85.7
25	17				774	9767	13080	68.5	87.8
26	18				714	9785	14015	69.6	90.4

* Solids S.G.-3.32

Table No. 8 - Bioleach Daily operating Data

Date	Oxygen		pH	Temp. (C)	EMF (mV)	Liquor S.G. (g/mL)	ASSAYS, mg/L						Py Oxidation Rate g/L/day	Comments
	Dissolved (mg/L)	Uptake (mg/L/hr)					Fe 2+	Fe 3+	FeTotal	As	Fe(AsPy)	Fe (Py)		
88/12/09	-	-	1.96	30	425	-	4280	302	4782	362	270	4512		100 ml Thiobacillus sp. + nutrients added
88/12/12	-	-	2.11	30	470	-	4784	1816	6600	441	329	6271		
88/12/13	-	-	2.23	32	460	1.014	4621	779	5400	222	166	5234		
88/12/14	-	-	2.12	34	435	1.012	5388	312	5700	257	192	5508		
88/12/15	-	-	2.12	34	490	1.015	7158	342	7500	292	218	7282	1.77	
•			-	34	-	-	4956	2644	7600	-	-	-	-	after increasing mixer speed and aeration
88/12/16	5.90	90	1.15	34	505	1.024	3251	4847	8098	908	677	7421	0.96	remove 1L pulp for Cyn-1
88/12/17	5.95	55	1.20	34	505	1.026	3646	4951	8597	-	-	-	-	
88/12/18	4.92	39	1.35	35	505	1.035	4672	6288	10960	-	-	-	-	
88/12/19	4.92	34	1.35	35	505	1.040	4225	6115	10341	-	-	-	-	300 ml innoculation from beaker + 100 ml from Dickenson stored culture
•														
88/12/20	8.04	19	1.39	28	505	1.042	4649	3699	12140	1950	1454	10686	0.86	
88/12/21	5.71	50	1.25	34	505	1.056	3646	12306	15952	3340	2490	13462	1.14	
88/12/22	5.87	68	1.34	35	565	1.052	1025	12192	13217	-	-	-	-	remove 1L pulp for Cyn-2
88/12/23	5.79	72	1.29	35	645	1.046	285	14470	14755	-	-	-	-	remove 1L pulp for Cyn-3
88/12/24	5.84	80	1.34	34	700	1.054	<10	13600	13610	6030	4496	9114	0.36	remove 1.4 L pulp for Cyn-4
88/12/25	6.80	66	1.28	33	<700	1.046	<10	13070	13080	4250	3169	9911	0.40	remove 2L pulp for Cyn-5
88/12/26	7.59	68	1.28	35	<700	1.052	<10	13260	13270	4330	3229	10041	0.38	remove 3L pulp for Cyn-6
88/12/28	6.50	64	1.30	35	<700	1.052	<10	13400	13410	4380	3266	9030	0.25	

TEST NO. CYN-1

Purpose: To investigate the recovery of gold from the bioleached residue by cyanidation.

Procedure: The residue was washed with weak H_2SO_4, filtered, repulped with water and washed again. The washed residue was placed in a small bottle, repulped with water and neutralized with lime for 2.5 hours prior to cyanidation.

Feed: 100 g (wet) residue (8 days)

Solution Volume: 200 mL Pulp Density 33% Solids

Solution Composition: 2.0 g/L NaCN

pH Range: 10.5-11 with $Ca(OH)_2$

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
0-1	-	0.76	-	0.58	-	-	-	-	11.1-9.6
1-2.5	-	0.41	-	0.31	-	-	-	-	11.2-11.0
0-2	0.42	-	0.40	-	0.16	-	0.24	-	11.0-10.5
2-19	0.25	0.23	0.24	0.17	0.12	-	0.28	-	11.1-10.5
19-24	0.29	0.24	0.28	0.18	0.38	0.03	0.02	-	11.2-11.1
Total	0.96	1.64	0.92	1.24	0.38	0.03	0.54	1.21	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 9.20 kg/t CaO:20.61 kg/t

Metallurgical Results

Product	Amount	Assays %, mg/L, g/t				% Distribution Au
		Au	SO_4	Fe	As	
1. Preg + Wash Solution	610 mL	0.73	-	7.31	3.10	40.5
2. Residue	58.7 g	11.13	1.74	9.86	4.82	59.5
Head (Calc)	58.7 g	18.17	-	9.86	4.82	100.0

TEST NO. CYN-2

Purpose:	To investigate the extraction of gold by cyanide leaching from the Bio-leach residue of Test 19
Procedure:	Transfer the washed residue to a 1 liter bottle and dilute to 33% solids. Neutralize with lime and maintain pH 11 for 2 hours. Add NaCN and carbon and continue agitation for a further 24 hours.
Feed:	60 g washed residue from Bio-leach (14 days)
Solution Volume:	150 mL Pulp Density 30% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with $Ca(OH)_2$
	Carbon: 10 g/L (1.5 g)

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Neutralize									
0-1	-	2.25	-	1.69	-	-	-	-	11.0-10.4
1-2	-	0.20	-	0.15	-	-	-	-	11.1-10.9
Cyanidation									
0-5	0.34	-	0.32	-	0.10	-	0.22	-	11.0-10.4
5-24	0.23	0.15	0.22	-	0.07	0.02	0.25	1.83	-10.5
Total	0.57	2.6	0.54	1.84	0.07	0.02	0.47	1.83	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 6.49 kg/t CaO: 25.2 kg/t

Test No. Cyn-2

Metallurgical Results

Product	Amount		Assay: g/t,mg/L Au	% Distribution Au
1. Carbon	1.7	g	685	67.3
2. Barren &Wash Sol'n	286	ml	0.030	0.5
3. Residue	72.4	g	7.68	32.2
Head (calc)	72.4	g	23.88	100.0

TEST NO. CYN-3

Purpose:	To investigate the extraction of gold by cyanide leaching from the Bio-leach residue of Test 19
Procedure:	As for Test No. CYN-2
Feed:	50 g of washed residue from Bio-Leach (15 days)
Solution Volume:	110 mL Pulp Density 33% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with $Ca(OH)_2$
Carbon:	10 g/L (1.1 g)

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Neutralize									
0-1	-	1.00	-	0.75	-	-	-	-	11.7-11.4
1-2	-	-	-	-	-	-	-	-	-11.1
Cyanidation									
0-5	0.24	-	0.23	-	0.08	-	0.15	-	11.2-10.5
5-24	0.16	0.10	0.15	0.08	0.06	0.03	0.17	0.80	-10.7
Total	0.40	1.10	0.38	0.83	0.06	0.03	0.32	0.80	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 6.78 kg/t CaO: 16.9 kg/t

Test No. Cyn-3

Metallurgical Results

Product	Amount		Assay: g/t,mg/L Au	% Distribution Au
1. Carbon	1.3	g	885	83.3
2. Barren &Wash Sol'n	275	ml	0.036	0.7
3. Residue	47.2	g	4.68	16.0
Head (calc)	47.2	g	29.26	100.0

TEST NO. CYN-4

Purpose:	To investigate the extraction of gold by cyanide leaching from the Bio-leach residue of Test 19
Procedure:	As for Test No. CYN-2
Feed:	75 g of washed residue from Bio-Leach (16 days)
Solution Volume:	150 mL Pulp Density 33% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with $Ca(OH)_2$
Carbon:	10 g/L (1.5 g)

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Neutralize									
0-1	-	0.75	-	0.56	-	-	-	-	11.1-10.3
1-2	-	0.20	-	0.15	-	-	-	-	11.1-11.0
Cyanidation									
0-5	0.32	-	0.30	-	0.12	-	0.18	-	11.1-10.5
5-24	0.19	0.12	0.18	0.09	0.05	0.02	0.25	0.78	-10.6
Total	0.51	1.07	0.48	0.80	0.05	0.02	0.43	0.78	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 6.84 kg/t CaO: 12.4 kg/t

Test No. Cyn-4

Metallurgical Results

Product	Amount	Assay: g/t,mg/L Au	% Distribution Au
1. Carbon	1.6 g	967	84.7
2. Barren &Wash Sol'n	300 ml	0.058	1.0
3. Residue	62.9 g	4.17	14.4
Head (calc)	62.9 g	29.04	100.0

TEST NO. CYN-5

Purpose:	To investigate the extraction of gold by cyanide leaching from the Bio-leach residue of Test 19
Procedure:	As for Test No. CYN-2
Feed:	56 g of washed residue from Bio-Leach (17 days)
Solution Volume:	112 mL Pulp Density 33% Solids
Solution Composition:	2.0 g/L NaCN
pH Range:	10.5-11 with $Ca(OH)_2$
Carbon:	10 g/L (1.1 g)

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Neutralize									
0-1	-	0.70	-	0.53	-	-	-	-	11.0-10.1
1-2	-	0.30	-	0.22	-	-	-	-	11.2-11.0
Cyanidation									
0-5	0.23	-	0.22	-	0.08	-	0.14	-	11.1-10.4
5-24	0.15	0.12	0.14	0.09	0.03	0.01	0.19	0.82	-10.6
Total	0.38	1.12	0.36	0.84	0.03	0.01	0.33	0.82	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 6.37 kg/t CaO: 15.83 kg/t

Test No. Cyn-5

Metallurgical Results

Product	Amount		Assay: g/t,mg/L Au	% Distribution Au
1. Carbon	2.4	g	580	87.3
2. Barren &Wash Sol'n	250	ml	0.031	0.5
3. Residue	51.8	g	3.75	12.2
Head (calc)	51.8	g	30.77	100.0

TEST NO. CYN-6

Purpose: To investigate the extraction of gold by cyanide leaching from the Bio-leach residue of Test 19

Procedure: As for Test No. CYN-2

Feed: 70 g of washed residue from Bio-Leach (158 days)

Solution Volume: 140 mL Pulp Density 33% Solids

Solution Composition: 2.0 g/L NaCN

pH Range: 10.5-11 with $Ca(OH)_2$

Carbon: 10 g/L (1.4 g)

Time Hours	Added, grams				Residual		Consumed		pH
	Actual		Equivalent		Grams		Grams		
	NaCN	$Ca(OH)_2$	NaCN	CaO	NaCN	CaO	NaCN	CaO	
Neutralize									
0-1	-	1.10	-	0.83	-	-	-	-	11.1-11.0
1-2	-	-	-	-	-	-	-	-	-10.9
Cyanidation									
0-5	0.30	-	0.28	-	0.10	-	0.18	-	11.0-10.4
5-24	0.19	0.15	0.18	0.11	0.07	0.02	0.21	0.92	-10.6
Total	0.49	1.25	0.46	0.94	0.07	0.02	0.39	0.92	-

Reagent Consumption (kg/t of cyanide feed) NaCN: 7.53 kg/t CaO: 17.76 kg/t

Test No. Cyn-6

Metallurgical Results

Product	Amount		Assay: g/t,mg/L Au	% Distribution Au
1. Carbon	1.6	g	948	89.5
2. Barren &Wash Sol'n	256	ml	0.062	0.9
3. Residue	59.5	g	2.72	9.6
Head (calc)	59.5	g	28.48	100.0

TEST NO. 20

Purpose: To perform a neutralization test on the 500 mL of Bioleach liquor.

Procedure: Mix and neutralize with 20% lime slurry to pH 5 for 1 hour and pH 11 for 1 hour. Filter and save all products.

Data: Initial pH - 1.3

62 mLs of $Ca(OH)_2$ slurry added to neutralize to pH 5

62 mL x 20% slurry = 12.4 g $Ca(OH)_2$

= 24.8 kg/t $Ca(OH)_2$

39 mLs of $Ca(OH)_2$ slurry added to neutralize to pH 11

39 mLs x 20% slurry = 7.8 g $CaOH_2$

= 15.6 kg/t $Ca(OH)_2$

Wet Weight of Precipitate = 166.8 g

Dry Weight of Precipitate = 58.3 g

LAKEFIELD RESEARCH
A Division of Falconbridge Limited
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